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Design Manual
Neutralization of Acid Mine Drainage

Penn Environmental Consultants, Inc.
Pittsburgh, PA

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Design Manual:

Neutralization of Acid Mine Drainage

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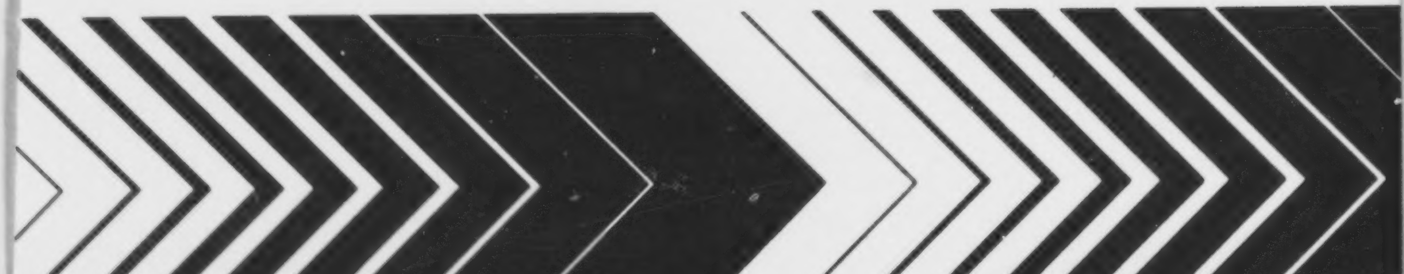

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This manual was prepared to assist designers and operators of mine drainage treatment plants in the selection of processes, equipment, and procedures. Included is a review of the most popular neutralizing agents and the methods used to handle, prepare, and feed these alkalis. Also, a detailed engineering explanation of the various processes applicable to treatment are presented.

The practical methods of sludge dewatering and disposal are explained along with modes of operation to improve solids content of the final volume. Techniques for lagooning and closure of such facilities are also discussed.

Examples of two treatment facility designs are included delineating general equipment specifications. Concluding the manual is a cost estimating section with example estimates based on the two sample designs. A general cost curve is also presented to allow for rapid estimating of treatment facilities based on flow.

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DESIGN MANUAL

NEUTRALIZATION OF ACID MINE DRAINAGE

U.S. ENVIRONMENTAL PROTECTION AGENCY

Office of Research and Development

Industrial Environmental Research Laboratory

NOTICE

This document has been reviewed in accordance with U.S. Environmental Protection Agency policy and approved for publication. Mention of trade names or commercial products does not constitute endorsement or recommendation for use.

FOREWORD

When energy and material resources are extracted, processed, converted, and used, the related pollutional impacts on our environment and even on our health often require that new and increasingly more efficient pollution control methods be used. The Industrial Environmental Research Laboratory - Cincinnati (IERL-Ci) assists in developing and demonstrating new and improved methodologies that will meet these needs both efficiently and economically.

This report provides specific design suggestions for neutralization systems for acid mine drainage treatment. It details step-by-step procedures, advantages and disadvantages, and costs for a variety of mine drainage treatment options. It will be of primary use for industry and consultants and will be of interest to academia and regulatory agencies. For further information, please contact the Nonferrous Metals and Minerals Branch, Energy Pollution Control Division.

David G. Stephan
Director
Industrial Environmental Research Laboratory

ABSTRACT

This manual was prepared to assist designers and operators of mine drainage treatment plants in the selection of processes, equipment, and procedures. Included is a review of the most popular neutralizing agents and the methods used to handle, prepare, and feed these alkalis. Also, a detailed engineering explanation of the various processes applicable to treatment are presented.

Examples of two treatment facility designs are included, delineating general equipment specifications and cost breakdowns.

The practical methods of sludge dewatering and disposal are explained, along with modes of operation to improve solids content of the final volume. Techniques for lagooning and closure of such facilities are also discussed.

Concluding the manual is a cost curve for the installation of treatment plants of various sizes. This curve will allow designers to derive an estimated budget number for capital expenditures.

This report was submitted in fulfillment of Contract No. 68-03-2599 by Penn Environmental Consultants, Inc., under the sponsorship of the U.S. Environmental Protection Agency. It covers the period from September 1978 to May 1981.

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LIST OF CONVERSIONS

MULTIPLY (ENGLISH UNITS)		by		TO OBTAIN (METRIC UNITS)
English Unit	Abbreviation	Conversion	Abbreviation	Metric Unit
acres	ac	0.405	ha	hectares
acre-feet	ac-ft	1,233.5	m^3	cubic meters
British Thermal Units	BTU	0.252	kg cal	kilogram-calories
British Thermal Units/ pound	BTU/lb	0.555	kg cal/kg	kilogram-calories/kilogram
cubic feet	ft^3	0.028	m^3	cubic meters
cubic feet	ft^3	28.32	l	liters
cubic feet/minute	ft^3/min	0.028	m^3/min	cubic meters/minute
cubic feet/second	ft^3/s	1.7	m^3/min	cubic meters/minute
cubic inches	in^3	16.39	cm^3	cubic centimeters
cubic yards	yd^3	0.76456	m^3	cubic meters
degrees Fahrenheit	°F	0.555 (°F-32)[a]	°C	degrees Celsius
feet	ft	0.3048	m	meters
foot-pounds	ft-lb	0.13825	kg-m	kilogram-meters
flask of mercury	(76.5 lb)	34.73[a]	kg Hg	kilograms of mercury
gallons	gal	0.003785	m^3	cubic meters

LIST OF CONVERSIONS (continued)

MULTIPLY (ENGLISH UNITS)		by		TO OBTAIN (METRIC UNITS)
English Unit	Abbreviation	Conversion	Abbreviation	Metric Unit
gallons	gal	3.785	l	liters
gallons/day	gal/d	0.003785	m^3/d	cubic meters/day
gallons/minute	gal/min	0.0631	l/s	liters/second
horsepower	hp	0.7457	kW	kilowatts
inches	in	2.54	cm	centimeters
inches of mercury	in Hg	0.03342	atm	atmospheres
miles (statute)	mi	1.609	km	kilometers
million gallons/day	Mgal/d	3,785[a]	m^3/d	cubic meters/day
ounces (troy)	troy oz	31.10348	g	grams
pounds	lb	0.454	kg	kilograms
pounds/cubic foot	lb/ft^3	16.02	kg/m^3	kilograms/cubic meter
pounds/gallon	lb/gal	119.8	g/l	grams/liter
pounds/square inch (gauge)	lb/in^2g	$(0.06805\ lb/in^2g)$[a]	atm	atmospheres (absolute)
pounds/square inch (gauge)	lb/in^2g	5.1715	cm Hg	centimeters of mercury

LIST OF CONVERSIONS (continued)

MULTIPLY (ENGLISH UNITS)		by	TO OBTAIN (METRIC UNITS)	
English Unit	Abbreviation	Conversion	Abbreviation	Metric Unit
pounds/square inch (gauge)	lb/in^{2}g	0.0703	kg/cm^2	kilograms/square centimeter
square feet	ft^2	0.0929	m^2	square meters
square inches	in^2	6.452	cm^2	square centimeters
tons (short)	ton	0.907	kkg	metric tons (1,000 kilograms)
tons (long)	long ton	1.016	kkg	metric tons (1,000 kilograms)
yards	yd	0.9144	m	meters

---

[a] Actual conversion; not a multiplier.

CHAPTER 1

INTRODUCTION

1.1 Background

Acid mine drainages (AMD) are not new problems to the coal industry or to many other mining industries for that matter. One need drive only a few miles through the coal areas of Kentucky, West Virginia, and Pennsylvania, for example, to see the yellow-stained bottoms of streams that contain few living organisms.

Acid discharges existed long before the mining of coal began. The acid is formed when pyrite, which is iron sulfide, is exposed to oxygen and water. The pyrite oxidizes to form a weak solution of sulfuric acid. As the solution of sulfuric acid passes over the varieties of rock strata surrounding the pyrite, it dissolves such metals as iron, aluminum, manganese, calcium, magnesium, sodium, and possibly some trace metals such as arsenic, selenium, beryllium, nickel, zinc, and others.

Unfortunately, pyrite occurs naturally in close proximity to the coal seams. Thus, the mining of coal exposes vast quantities of pyritic material to water and oxygen and greatly accelerates the natural oxidation processes, resulting in the significant production of acid mine drainage.

Regulation of the concentration of certain chemical parameters in the discharges is now a way of life for the mine operator. Current new-source discharge guidelines, as proposed in the January 13, 1981, Federal Register (Vol. 46, No. 8), are shown in Table 1-1 for point-source acid mine drainage discharges. Depending upon the receiving stream's quality and flow, the limits may be further restricted by State permits.

Iron, aluminum, and manganese are acid-soluble, so merely neutralizing the water (increasing the pH) will precipitate these ions. This is not so easy as it sounds, however, because several factors complicate the precipitation. First, iron can exist in two forms in acid mine drainage; i.e., ferrous (unoxidized) and ferric (oxidized). The ferric (Fe^{3+}) form will begin to precipitate around pH 4.0, forming ferric hydroxide or more complicated oxy-hydroxides; this is the yellowboy common to stream beds in coal country. The ferrous (Fe^{2+}) form begins to precipitate at about pH 8.0 and forms a blue-green hydroxide. In fact, an easy test for significant ferrous iron concentration in AMD is to sprinkle lime into the drainage, and if the blue-green hue develops as the lime dissolves in the water, ferrous iron is present. It is usually advantageous to oxidize the ferrous iron to the ferric state rather than to rely upon ferrous precipitation at high pH's. This

TABLE 1-1

POINT-SOURCE DISCHARGE LIMITATIONS
FOR ACID MINE DRAINAGE
(NEW-SOURCE PERFORMANCE STANDARDS)

Effluent Characteristic	Effluent Limitations Maximum for Any 1 Day mg/l	 Average of Daily Values for 30 Consecutive Days Shall Not Exceed mg/l
Iron, total	7.0	3.5
Manganese, total	4.0	2.0
Total suspended solids (TSS)	70.0	35.0
pH	Within the range 6.0-9.0	--

oxidation is accomplished by increasing the pH above 7.0 and inducing air into the water to provide oxygen. The oxidation rate of ferrous iron is strongly pH-dependent and proceeds extremely slowly below pH 6.0.

Ferric iron and aluminum both begin to precipitate around pH 5.0. At extremely high pH's (above pH 10.0), the aluminum may tend to redissolve. A pH above 8.0 is necessary to precipitate manganese to achieve required effluent levels.

Although increasing the pH can remove all of the elements from solution as required by the regulations, many of the flocs (precipitates) that form, and especially iron flocs, are quite light and tend to remain suspended rather than settle. It is probably more difficult in treatment situations to remove the precipitates (flocs or sludge) from suspension than to increase the pH, oxidize, and precipitate them in the first place. The settling process is monitored by the total suspended solids (TSS) regulation in the guideline limitations (Table 1-1). All the regulated flocs eventually settle; however, the trick is to settle them before they reach the outlet of the settling basin or clarifier and thus be in compliance with the discharge regulations.

1.2 Review of AMD Chemistry

A review of the basic chemical reactions involved in the production of AMD is helpful in understanding the rationale involved in the design of the unit processes and the overall treatment scheme.

Pyrite (FeS_2) and marcasite (also FeS_2) are both naturally occurring minerals associated with coal-bearing strata. Pyrite and marcasite differ in crystalline structure but have identical iron disulfide composition. Although both are AMD-producers, pyrite is more abundant and is thus commonly credited for AMD generation. Pyrite occurs in several forms, based upon the geological conditions at the time of formation. The framboidal form is the prolific acid producer and is characterized by a cluster of spheres of agglomerated minute pyrite crystals, approximately 25 μm (micrometers) diameter.

Iron disulfides (both pyrite and marcasite) react with water and oxygen to form ferrous sulfate and to release 2 mol (moles) of hydrogen ions (acid) (1).

$$FeS_2 + \frac{7}{2} O_2 + H_2O = Fe^{+2} + 2SO_4^{-2} + 2H^+ \qquad (1)$$

The ferrous iron will eventually use 1 mol of H^+ and oxidize to the more stable ferric form according to:

$$Fe^{+2} + \tfrac{1}{4}O_2 + H^+ = Fe^{+3} + \tfrac{1}{2}H_2O \qquad (2)$$

Above approximately pH 4.0, the ferric ion will hydrolyze (take on water) and precipitate, freeing 3 mol of H^+.

$$Fe^{+3} + 3H_2O = Fe(OH)_3 + 3H^+ \qquad (3)$$

Overall, 1 mol of pyrite will produce 4 mol of hydrogen ions: 2 mol from the initial oxidation of pyrite and 2 mol (net) from the combined iron oxidation of ferrous to ferric and the subsequent hydrolysis to ferric hydroxide. The 4 mol of H^+ are equivalent to 2 mol of sulfuric acid (H_2SO_4).

Thus, the acidity in the AMD entering a treatment plant is developed not only from the pyrite but also from the hydrolysis of the metals as they oxidize, hydrolyze, and precipitate. This fact is important to treatment plant design and operation because it governs the amount of alkali ultimately required for neutralization. For example, Equations 2 and 3 indicate that a net 2 mol of H^+ (equivalent to 1 mol of H_2SO_4) are produced from oxidizing and precipitating 1 mol of ferrous iron. Using an inlet ferrous iron concentration of 100 mg/l, 62 mg/l (0.6 lb/1,000 gal) of hydrated lime would ideally be required to neutralize the additional acidity from the ferrous iron oxidation/precipitation. Analytical methods for determining acidity in AMD should, and normally do, include the addition of a strong oxidant such as hydrogen peroxide to oxidize all the metals, thus producing a measurement of total acidity, which includes acidity resulting from oxidation of all metals and the hydrolysis/precipitation of those metals insoluble at the end point of the acidity titration (normally between pH 7.3 and 8.3).

This total acidity value is normally expressed in terms of milligrams per liter (mg/l) of calcium carbonate ($CaCO_3$), meaning ideally that if the number of milligrams of $CaCO_3$ specified were added to 1 l of the AMD, the pH of the AMD would ultimately increase to the titration end point (normally between pH 7.3 and 8.3). It is important to recognize that the operator may, and probably will, operate the system at a different pH level than the end point of the acidity titration; thus, the theoretical alkali requirement for treatment may differ from that predicted by the acidity determinations.

1.3 Acidity

Acidity is a major factor to be considered in system design because it strongly influences the choice of alkali. To calculate the theoretical "ideal" amount of alkali required to neutralize a given acidity concentration, the alkalis must be converted to an equivalent basis; i.e., expressed as calcium carbonate. The equivalent weight is the molecular weight divided by the valence of the dissociated ions. For example, the molecular weight of $CaCO_3$ is 100.09 and the valence of Ca is 2; therefore, the equivalent weight is 100.09 ÷ 2 = 50 (rounded). Similarly, sodium hydroxide (NaOH) has a molecular weight of 39.99, the valence of Na is 1, and the equivalent weight is 39.99 ÷ 1 = 40. Soda ash (Na_2CO_3) dissociates into $2(Na^{+1})$ and $(CO_3)^2$; thus, the equivalent weight is 105.99 ÷ 2 = 53. The lower the equivalent weight, the more powerful the alkali, because less reagent is necessary to provide an equivalent neutralization capability. For example, if 50 mg of $CaCO_3$ were required to neutralize 1 l of AMD with an acidity of 50 mg/l, only 28 mg of quicklime or CaO (see Table 1-2) would be necessary to provide an equal amount of neutralization.

Once the ideal alkali requirement has been determined, it is necessary to compensate for the generally inefficient utilization of the reagent in the actual treatment process. For example, hydrated lime utilization efficiencies are typically near 70%, thus requiring 1.4 times the ideal theoretical amount to accomplish neutralization; limestone efficiencies are below 50%, thus requiring over two times the ideal quantity; and, sodium hydroxide efficiencies are above 90%, requiring only 1.1 times the theoretical reagent quantity.

1.4 Ion Solubility and pH

The metal ions normally present in AMD are typically relatively insoluble in alkaline environments, and therefore can be precipitated as hydroxides by increasing the pH. The theoretical solubilities, determined by measurements of each individual ion dissolved in distilled water, are illustrated in Figure 1-1. Actual mine waters involve complex interactions and result in shifts of the curves shown in Figure 1-1; however, the general trends remain the same. Manganese, for example, can generally be precipitated at pH's slightly above 8.0, probably because of coprecipitation with iron. In the infrequent situation that manganese cannot be removed within the pH 6.0-9.0 requirements of the New-Source Performance Standards (Table 1-1), the regulations allow elevating the pH slightly above 9.0 to achieve satisfactory manganese removal.

TABLE 1-2

ALKALI COMPARISON FOR TREATMENT OF ACID MINE DRAINAGE

Alkali	Formula	Molecular wt g	Equivalent wt g	Factor to Convert to $CaCO_3$ Equivalence
Calcium neutralizers				
Hydrated lime (calcium hydroxide)	$Ca(OH)_2$	74.10	37.05	1.35
Quicklime (calcium oxide)	CaO	56.08	28.04	1.78
Limestone (calcium carbonate)	$CaCO_3$	100.08	50.04	1.00
Magnesium neutralizers				
Dolomitic lime (magnesium hydroxide)	$Mg(OH)_2$	58.30	29.15	1.72
Sodium neutralizers				
Caustic soda (sodium hydroxide)	$NaOH$	39.99	39.99	1.25
Soda ash (sodium carbonate)	Na_2CO_3	105.99	53.00	0.94

1.5 References

1. Singer, P.C., and W. Stumm. Oxygenation of Ferrous Iron. Federal Water Pollution Control Administration Research Series 14010, Cincinnati, Ohio, June 1969.

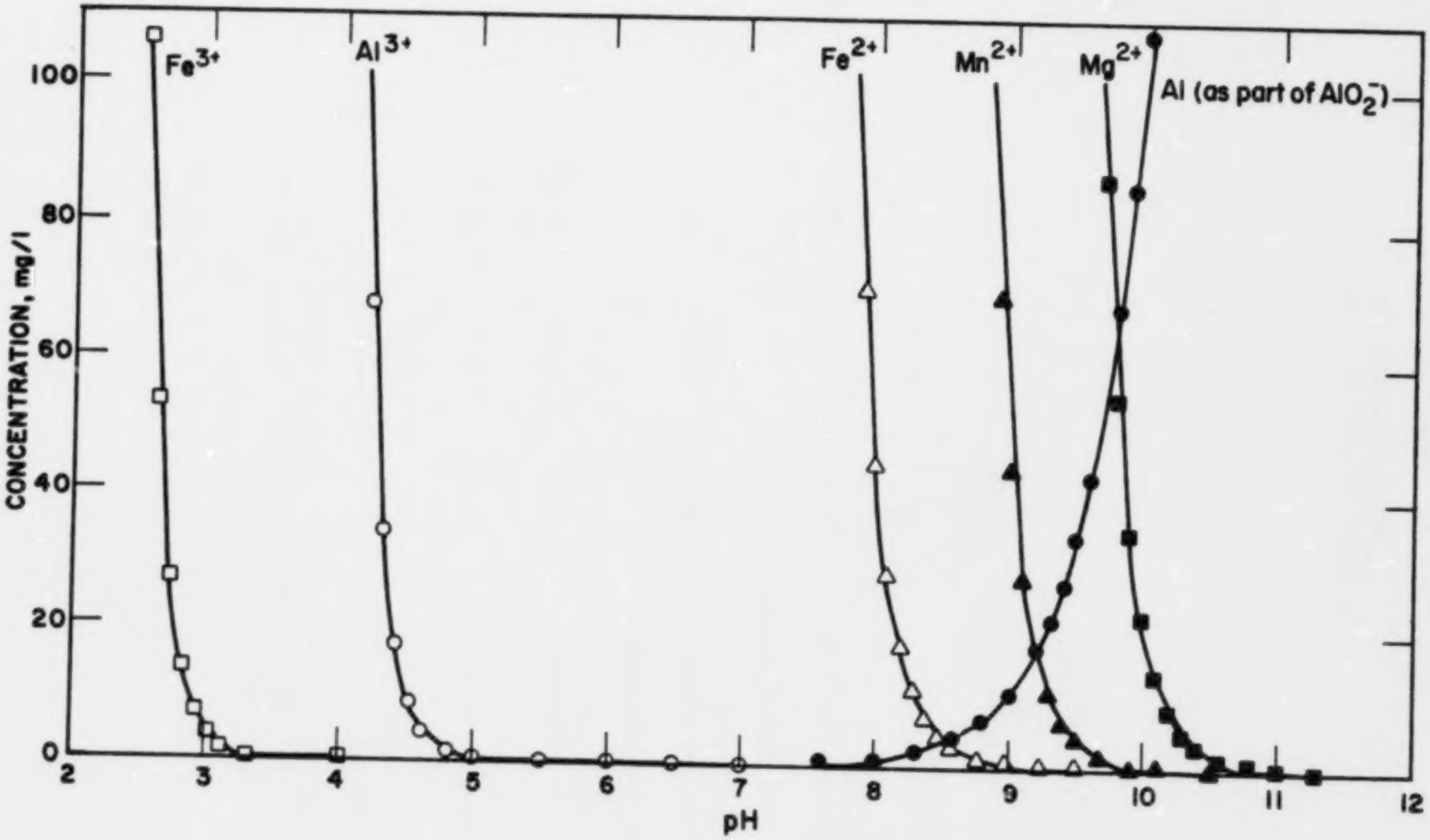

Figure 1-1. Theoretical solubilities of selected ions.

CHAPTER 2

GENERAL TREATMENT CONSIDERATIONS

2.1 Acid Mine Drainage Treatment Systems

As previously described, AMD is a dilute solution of sulfuric acid and iron sulfate with iron in the ferrous and/or ferric form. Its treatment consists of neutralization with a suitable alkali, oxidation of ferrous iron to the insoluble ferric form, and removal of the resulting metal precipitants by a sedimentation process. There is one basic process system for the treatment of AMD; however, there are many options available to the designer when evaluating each of the unit processes or subprocesses within the overall system. Figure 2-1 shows many of these options within the overall process flow sheet. Consequently, this manual has been developed in a format that discusses each unit operation (chemical feeding, mixing, sedimentation) of the process separately. A discussion of the various overall processes follows.

2.1.1 Conventional Lime Neutralization Process (2, 3, 4)

In the conventional lime neutralization process, each of the five basic treatment steps follows in normal sequence; i.e., equalization, neutralization (mixing), aeration, sedimentation, and sludge disposal. Flow is once-through and gravity systems are usually employed. A flow sheet for the typical system is shown in Figure 2-2.

To simplify the controls needed in the system and to minimize operator attendance, a constant flow with only small variations in quality is desirable. To accomplish this goal, the mine drainage is collected in large holding or equalization basins, or in large sumps within active portions of the mine. Such holding basins should have a capacity for storage of 2-3 days flow during shutdown periods. Normally, 12-24 hours flow is maintained in the holding basin to equalize flow and quality to the treatment facility. From the holding basin, the mine drainage either flows by gravity or is pumped to the treatment plant. Since most mines are in rural areas, both the holding and settling basins are usually surface impoundments of earthen construction. (Earthen pond design is discussed in Chapter 6.)

Lime is used as the alkali in practically all large-volume AMD treatment plants. The selection between quicklime and the hydrate is determined by availability, cost, or personal preference. This process is discussed in detail in Chapter 3, as are the several methods by which the lime can be fed into the mixing or neutralization unit. The use of other alkalis is also discussed in Chapter 3.

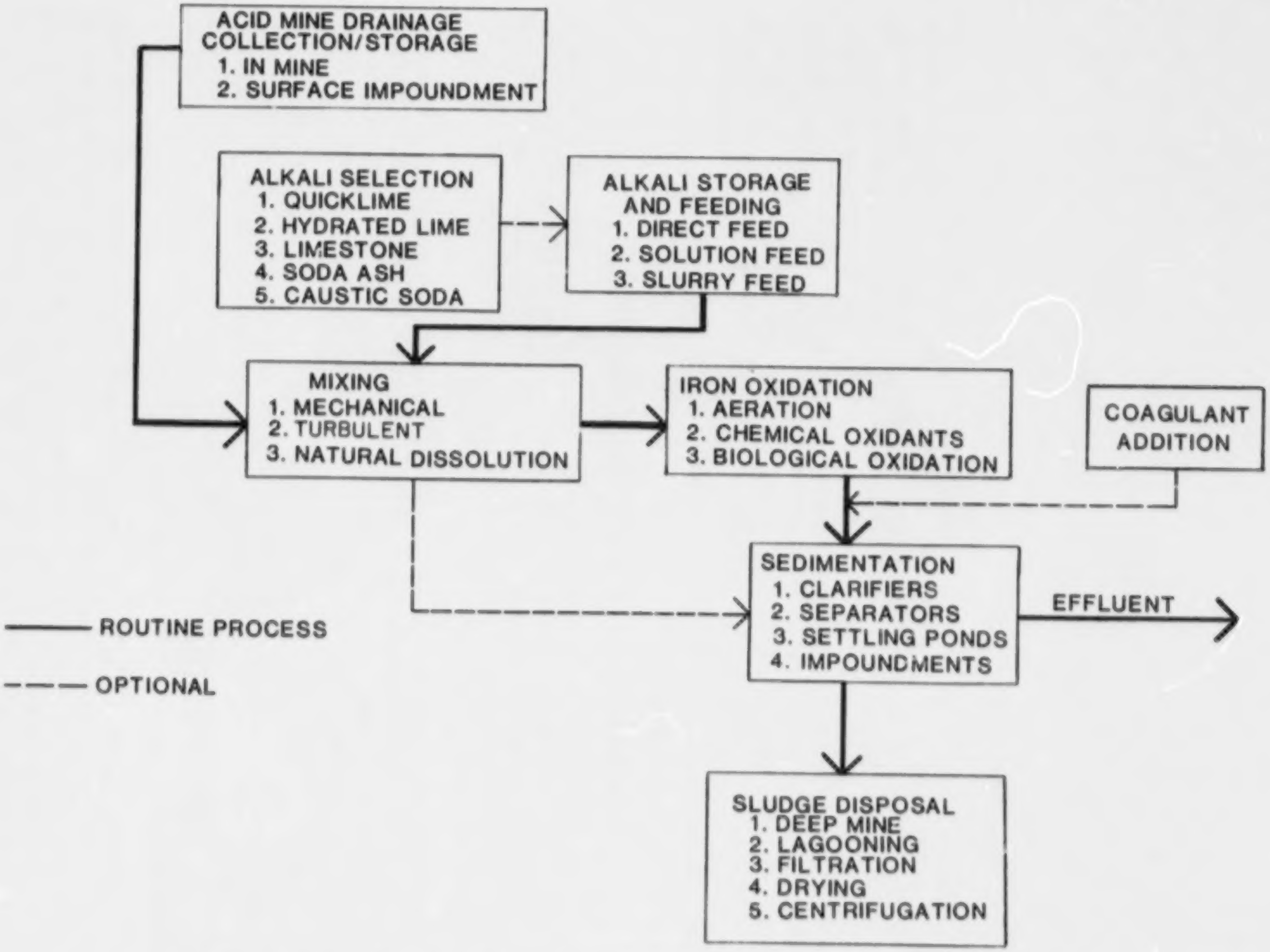
ACID MINE DRAINAGE COLLECTION/STORAGE
1. IN MINE
2. SURFACE IMPOUNDMENT
ALKALI SELECTION
1. QUICKLIME
2. HYDRATED LIME
3. LIMESTONE
4. SODA ASH
5. CAUSTIC SODA
ALKALI STORAGE AND FEEDING
1. DIRECT FEED
2. SOLUTION FEED
3. SLURRY FEED
MIXING
1. MECHANICAL
2. TURBULENT
3. NATURAL DISSOLUTION
IRON OXIDATION
1. AERATION
2. CHEMICAL OXIDANTS
3. BIOLOGICAL OXIDATION
COAGULANT ADDITION
SEDIMENTATION
1. CLARIFIERS
2. SEPARATORS
3. SETTLING PONDS
4. IMPOUNDMENTS
EFFLUENT
SLUDGE DISPOSAL
1. DEEP MINE
2. LAGOONING
3. FILTRATION
4. DRYING
5. CENTRIFUGATION
ROUTINE PROCESS
OPTIONAL

Figure 2-1. Process flow sheet for treatment of acid mine drainage.

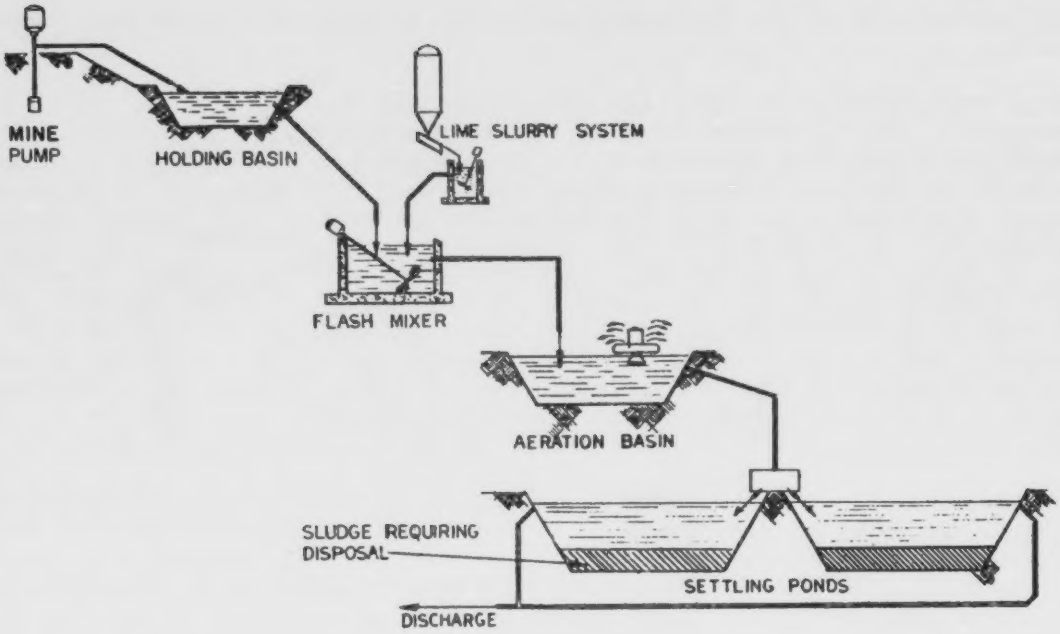

Figure 2-2. Conventional lime neutralization process.

Aeration is a straightforward process for oxidizing ferrous iron to the less soluble ferric form. Ferrous iron is much more soluble than the ferric form, with minimum solubility occurring in the pH range of 9.3-12.0 (1). Ferric iron, on the other hand, is much less soluble and begins to precipitate as a hydroxide at a pH of 4.0, with minimum solubility occurring about pH 8.0. Obviously there is an economic advantage in removing iron in the ferric form at the lower pH. Less lime is required for neutralization to the pH level needed to maintain minimum iron solubility (8.0 vs. 12.0).

The forced oxidation of ferrous iron is usually included in AMD treatment. This oxidation is pH-dependent, with the reaction proceeding rapidly at a pH above 8.0. With the pH requirement satisified, iron oxidation becomes dependent upon the availability of oxygen. This oxidation reaction is expressed by Equation 2, previously discussed in Chapter 1. The theoretical oxygen requirement is one unit weight for each seven weights of ferrous iron to be oxidized.

It is important to point out that most existing aeration units do not have sufficient hydraulic detention capacity; this may be one major cause of effluent compliance problems with iron. Chapter 5 discusses the design requirements for iron oxidation and the aeration equipment available to accomplish this task. Other methods for iron oxidation are also included in Chapter 5. It is worth noting that in a few cases there are advantages in precipitating iron as ferrous hydroxide because of peculiarities in water chemistry.

Once the drainage has been neutralized and the ferrous iron oxidized, if necessary, the subsequent step in the treatment process is sedimentation. Settling of the iron hydroxide and other suspended solids is commonly accomplished in earthen settling basins. These must have at least 12 hours of clear water detention above the sludge storage zone to meet minimum design requirements in Pennsylvania (5). Other design considerations are discussed in Chapter 6. When using small settling basins with 12-48 hours detention and minimal sludge storage capacity, two units operated in parallel are highly recommended to allow sufficient time for sludge removal without interrupting the treatment process. If treatment plant site conditions allow, large impoundments that provide many years of sludge storage can be advantageous. Chapter 6 includes design information for other sedimentation units; e.g., mechanical clarifiers, thickeners, and tilted-plate separators.

A necessary part of the treatment process is the need for adequate planning for sludge handling and disposal. This will be a significant part of both the construction and operating costs of the system. Without proper planning and process selection in this area, day-to-day treatment plant operation can become a considerable and overly expensive problem.

Even though many mine drainage treatment facilities have been in operation for 10 years or so, the handling and disposal of the sludge produced continues to be a problem. The simplest method for final disposal is to pump the sludge into abandoned deep mines. Although this practice is widespread, the overall environmental effects of this disposal method have yet to be determined. Unfortunately, this practice cannot be used at all facilities. Another method that has been used successfully is lagooning, where the sludge thickens naturally. Eventually the sludge must be disposed of in a more satisfactory manner, such as burial in a surface mine reclamation project.

There are other methods for sludge dewatering that are used infrequently but can be considered by the designers. Among these are drying beds, vacuum and pressure filtration, and centrifugation. Each of these is discussed in Chapter 7, Sludge Dewatering and Disposal.

2.1.2 High-Density Sludge Process (4, 6)

Variations on the conventional lime neutralization process previously discussed are the sludge recirculation processes that can be utilized to achieve better reactivity of the lime and produce smaller volumes of sludge containing higher solids. One such procedure is the High-Density Sludge Process developed in 1970 by the Bethlehem Steel Corporation. This process uses lime for neutralization and can produce a dense sludge that reduces its volume significantly more than the conventional process. The process is based on a high sludge recirculation rate within the system, where the optimum ratio of solids recirculated to solids removed is in the range of 20:1 to 30:1. The sludge is returned to a reactor vessel where the lime slurry is added. This point of alkali introduction is peculiar to the Bethlehem system. The slurry is then mixed with the AMD in a neutralization reactor, where aeration is provided for oxidation of ferrous iron. The process flow sheet is shown in Figure 2-3.

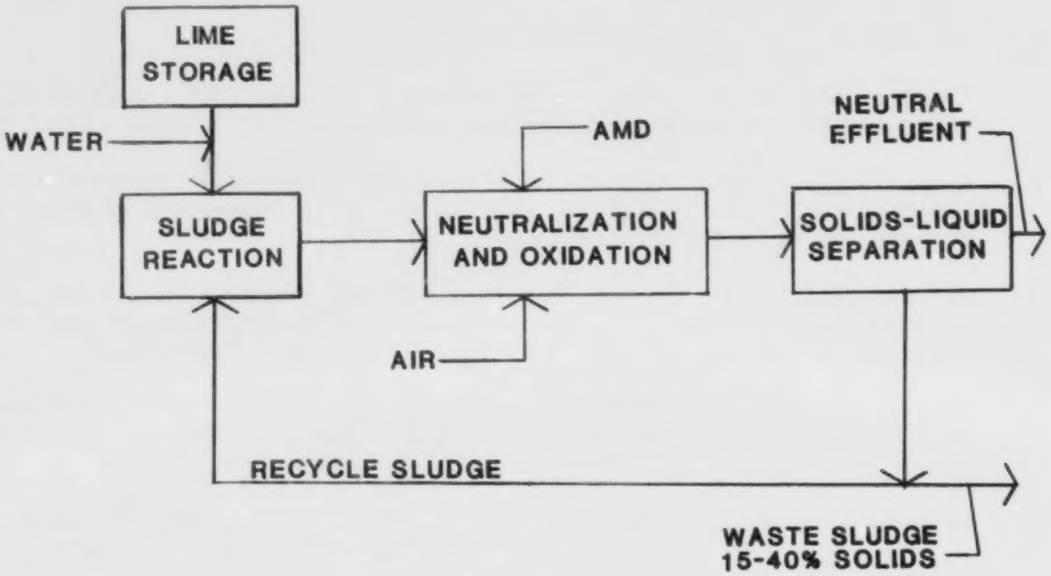
LIME STORAGE
WATER
AMD
NEUTRAL EFFLUENT
SLUDGE REACTION
NEUTRALIZATION AND OXIDATION
SOLIDS-LIQUID SEPARATION
AIR
RECYCLE SLUDGE
WASTE SLUDGE 15-40% SOLIDS

Figure 2-3. High-density sludge treatment process.

Removal of the solids is accomplished in large mechanical thickeners. The achievement of high sludge solids, reported by Bethlehem to be as great as 50%, is dependent upon the ferrous-to-ferric iron ratio. If ferric iron dominates this ratio, sludge densities may be limited to 20% solids. The user is cautioned that certain aspects of this system may be covered by patents issued to the Bethlehem Steel Corporation, but other variations on the sludge recirculation processes are common practice in AMD treatment.

2.1.3 Other Treatment Processes

There can be numerous variations on the conventional process for treatment of AMD. Where acidity is the main problem and the flow is low, other alkalis such as soda ash, caustic soda, or limestone can be used. Portable caustic soda treatment units are common in surface mine operations. Limestone has been used in several applications for in-place treatment. These applications are discussed in more detail in Chapter 3.

In addition to the treatment of AMD to achieve a desired effluent quality for discharge, other methods of treatment are available to produce a product water of higher quality. These include reverse osmosis, ion exchange, and chemical softening; discussions of these methods are presented in Chapters 9, 10, and 11, respectively. Each of these processes can possibly be utilized to produce water acceptable for human consumption.

2.2 References

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2. Dorr Oliver, Inc. Operation Yellowboy-Mine Drainage Treatment Plants and Cost Evaluation. Report to the Pennsylvania Department of Mines and Mineral Industries, Coal Research Board, 1966.

3. Holland, C.T., J.L. Corsaro, and D.J. Ladish. Factors in The Design of an Acid Mine Drainage Treatment Plant. Second Symposium on Coal Mine Drainage Research, Mellon Institute, Pittsburgh, Pennsylvania, 1968.

4. Skelly and Loy and Penn Environmental Consultants, Inc. Processes, Procedures, and Methods to Control Pollution from Mining Activities. EPA-430/9-73-011, Washington, D.C., 1973.

5. Bureau of Water Quality Management. Mine Drainage Manual. 2nd ed. Department of Environmental Resources, Publication No. 12, Harrisburg, Pennsylvania, September 1973.

6. Haines, G.F., and P.D. Kostenbader. High-Density Sludge Process for Treating Acid Mine Drainage. Third Symposium on Coal Mine Drainage Research, Mellon Institute, Pittsburgh, Pennsylvania, 1970.

CHAPTER 3

CHEMICAL TREATMENT

3.1 Introduction

The chemical treatment of acid mine drainage has involved the use of practically every available neutralizing agent in either pilot or full-scale operation. Information concerning the application and results associated with the use of these many reagents is beyond the scope of this manual. Only the more practical and commonly used alkalis are included. Chemical and physical data for lime and its solutions are in Appendix A. Presented in the following chapter are cost comparisons and descriptions of the process systems for the handling, storage, and feeding of each alkali. Those alkalis included are quicklime, hydrated lime, limestone, caustic soda, and soda ash.

More than 90% of all facilities treating acid mine drainage utilize a form of lime. The main factors affecting the selection of any alkali are cost, suitability, reactivity, availability, ease of use, and sludge volume. It is important for the designer to evaluate these factors fully because each alkali requires significantly different equipment, which limits further changeover to another type. In addition, the alkali selected may affect the design of other processes in the overall system.

3.2 Lime

Lime is a general term that, by definition, encompasses only burned forms of limestone. The two forms of particular interest in AMD treatment are quicklime and hydrated lime. Carbonates, such as limestone, are frequently but erroneously referred to as "lime." For this and other reasons, limestone is described separately.

3.2.1 Quicklime

Quicklime (CaO) is a product resulting from the calcination of limestone. Limestone basically consists of 50%-90% calcium carbonate ($CaCO_3$). When limestone is burned in a kiln at a temperature of about 1,000°C (1,835°F), carbon dioxide (CO_2) gas is driven off and calcium oxide (CaO) or quicklime is produced. On the basis of its chemical analysis, quicklime may be divided into three classes:

1. high calcium quicklime - containing less than 5% magnesium oxide;
2. magnesium quicklime - containing 5%-35% magnesium oxide; and

3. dolomitic quicklime - containing 35%-40% magnesium oxide.

Quicklime is available in a number of standard sizes, but the following describes those most applicable to mine drainage treatment:

Ground lime - the product resulting from grinding the larger sized material and/or screening off the fine size. A typical size is the majority passing a #8 sieve and 40%-60% passing a #100 sieve.

Pulverized lime - the product resulting from a more intense grinding than is used to produce ground lime. A typical size is the majority passing a #20 sieve and 85%-95% passing a #100 sieve.

Quicklime is most often obtained in either bulk carloads or 18.2-Mg (20-ton) pneumatic trucks, and then transferred to a storage silo. To be used efficiently in mine drainage neutralization, the quicklime must be slaked (see 3.2.7, Quicklime Slaking Systems). The slaking process must be carefully controlled and requires daily attention. Quicklime, combined with a good slaking operation, offers a low unit cost per gram of acidity neutralized.

The primary disadvantages of a quicklime system are high capital investment for a slaker, grit removal, close operational control, and the danger to personnel of possible severe burns.

3.2.2 Hydrated Lime

Hydrated lime ($Ca(OH)_2$) is the most commonly used alkali for neutralizing acid mine drainage in existing treatment plants. It is preferred when lime consumption rates are low or when the cost for a slaking system is prohibitive.

Hydrated lime is air-classified to produce the fineness necessary to meet user requirements. Normal grades of hydrate used for chemical purposes will have 75%-95% passing a #200 sieve. Due to air classification, the commercial hydrate produced is purer than the quicklime because most of the impurities are rejected.

Hydrated lime is packaged in paper bags weighing 22.7 kg (50 lb) net. It is also available in bulk. Physical and chemical properties of hydrated lime can be found at the end of this chapter.

3.2.3 Lime Handling

Hydrated lime can be purchased in various quantities. Where daily requirements are small, less than 138 kg (300 lb) bagged lime may be preferred. The handling and storage operations are relatively simple, usually involving manual labor or mechanical equipment, depending on the volume involved.

Most often bulk lime, either quick or hydrated, is more efficient and economical to use. In such installations, the lime is delivered by truck and conveyed by mechanical or pneumatic systems into weathertight bins or silos. The factors determining the type of lime to use are discussed in the economics of this chapter.

3.2.3.1 Bagged Lime

Bagged lime is delivered loose or palletized in truck or box car, and generally handled by hand truck or forklift to storage. Unloading conveyors may be preferred for loose bags, particularly if there is a long distance between the unloading point and the storage area. For palletized shipments, forklift trucks are utilized to move the lime to storage or point of use, thus eliminating much manual labor. To facilitate bag dumping, the hopper located above the feeder can be fitted with a bag ripper and screen.

Bagged lime should be stored in dry areas. Hydrated lime is normally packaged in multiwall paper bags, but exposure to moisture will permeate the liner and cause caking. Bags of hydrated lime may be stacked 20 high without damaging the bottom bags. In dry storage, hydrate may be stored for periods of up to 1 year without encountering serious deterioration. Care should be exercised to use the material in the order it is received, rather than maintaining an inactive reserve that may not be consumed for several months or years.

3.2.3.2 Bulk Lime

Considerable savings can be realized by using bulk lime instead of bagged lime, not only in initial cost but also in reduced labor involved in handling. In additon, there are other advantages including faster loading, elimination of losses from broken bags and spillage, better housekeeping because modern handling systems are completely enclosed, and less dust hazard to employees.

Delivery of bulk lime to the treatment plant can be accomplished by a variety of truck or rail-car equipment. Truck transportation, particularly the blower truck, is generally the fastest, most common, and most economical way to handle bulk shipments.

The pneumatic truck has become popular because of its simplicity and speed of delivery and unloading. The lime is blown from the truck directly to silo storage via a 10-cm (4-in) pipeline that eliminates mechanical conveyors. The only extra equipment required is a safety release valve and dust collector mounted atop the silo to exhaust the conveying air. The release valve is important to accommodate the large volume of air, up to 36.4 m (1,300 ft), in the blower truck when the tank is empty. The valve can be a hinged manhole cover or a simple 20-cm (8-in) pipe with weighted, gasketed cover.

Today, the bag-type dust collector is used to filter air expelled during truck unloading. For hydrated lime, normally 0.09 m^2 (1.0 ft^2) cloth area/ 0.06 m^3 (2.0 ft^3) of air is recommended. Approximately 34.8 m^2 (375 ft^2) of cloth area are needed for a large truck rotary blower of 21 m^3 (750 ft^3)/min capacity. Additional cloth area of 9.3-27.9 m^2 (100-300 ft^2) may be justified, however, to accommodate the final cleanout period. These recommendations are conservative, and more commonly, vendors use an air-to-cloth ratio of 3:1. If more extensive dust collection is required, the design should tend to be conservative.

Blower trucks are available with compartment tank capacity varying from 19.6 to 36.4 m^3 (700-1,300 ft^3). (The latter delivers up to 18.2 kkg (20 tons) of hydrate and 21.8 kkg (24 tons) of pebble lime.) Air is provided by a trailer-mounted positive displacement rotary blower, which furnishes up to 21 m^3 (750 ft^3) of air/min. It is operated from a power takeoff or by a separate engine.

The pipe used for transferring lime to storage can be ordinary black iron pipe, but galvanized pipe is highly recommended because of its resistance to rust. All bends should be made with a minimum radius of 0.9-1.2 m (3-4 ft) to reduce wear and resistance to flow. The intake end of the pipe should be mounted vertically, with the bottom 1.2 m (4 ft) above ground level, and equipped with a quick-connect coupling for the rubber blowing hose on the truck.

The largest size of pebble lime that can be pumped efficiently from blower trucks is 3.2 cm (1.25 in), although a top size of 2.5 cm (1.0 in) is preferred. Pebble lime may be blown as much as 30.5 m (100 ft) vertically and 45.7 m (150 ft) in a combined vertical and horizontal run. For greater distances, the unloading time becomes excessively long.

Hydrated lime, however, can be blown readily to 91.4 m (300 ft) in a combined vertical and horizontal run. If lime has to be blown both vertically and horizontally, it is much easier to blow vertically first, then horizontally, rather than vice versa. Otherwise, there will be excessive wear at the 90° elbow because of the resistance provided by the column of lime in the vertical pipe section, and unloading will be delayed. It is advisable to use only one 90° turn in the piping system, or none if possible.

When hauled by rail, most bulk lime is shipped in covered hopper cars, the largest capable of hauling 90 kkg (100 tons) of quicklime or 45 kkg (50 tons) of hydrated lime. These cars have two to four compartments, each with a bottom discharge gate. Generally, the lime is discharged to an under-track hopper, then taken by screw conveyor or bucket elevator to plant storage, or an adapter is fastened to the discharge gate by clamps for pneumatic unloading. Air vibrators are usually attached to the hopper bottoms to facilitate unloading, particularly for hydrate. For protection against rain when not in use, the under-track hopper should be covered.

3.2.4 Storage of Lime

Since quicklime and hydrated lime are not corrosive, conventional steel or concrete bins and silos can be used for storage. The storage units, however, must be watertight. Of all the storage units used for lime, the steel silo with cone bottom is the most popular.

The variety of bin or silo designs for lime storage includes rectangular, square, hexagonal, and circular. The first three occupy less plant space because they utilize common walls and are generally easy to clean, but they have the disadvantage of causing material retention in the corners. With round silos, this accumulation or bridging is uncommon, although there is greater tendency for lime to arch than in rectangular bins. In either case, the storage units should always be designed with a hopper or conical base to facilitate discharge.

The decision to install one or more large storage units versus several small ones will depend on the individual plant and on such factors as daily lime requirements, type of delivery (rail or truck), and operation (continuous or intermittent). In any event, the total storage capacity should be at least twice the minimum truck or rail delivery to guarantee a lime supply while awaiting an order. Most mine drainage treatment plants have a steady lime demand and operate continuously. It may be prudent to provide at least 7 days storage capacity, and preferably 2-3 weeks.

The importance of ample storage capacity cannot be overemphasized, especially when it is the cheapest cost per unit volume in the overall treatment system. Then, too, by having large storage capacity, future plant expansion can be readily accommodated.

In designing a specific size of silo or bin, it is advisable to use a bulk density of 481 kg/m^3 (30 lb/ft^3) for hydrated lime, and 881 kg/m^3 (55 lb/ft^3) for quicklime. These density figures are subject to variation, depending on lime source, type of lime, particle size (pebble vs. pulverized), and gradation. With both products, these values are conservative; hence, bins will have additional capacity for denser limes. Figures 3-1 and 3-2 show typical silo installations.

3.2.5 Flowability

The flowability of lime varies from good, for pebble and granulated quicklime, to erratic, for pulverized quicklime and hydrated lime. Lime tends to absorb moisture readily, forming an adherent soft cake that can cause arching or bridging in storage. Hydrated lime, in particular, also tends to "rathole," because of its fluffy texture and possibly electrostatic charges. Then, after collapsing, the hydrate may become fluidized and flood the discharge.

Because of the inherent problems with lime flowability, several units have been developed to provide a uniform density from storage to the feeder. These include special considerations in bin construction, the use of external vibrators, internal antipacking and antiarching devices, and live bin bottoms.

There are various opinions as to the best bin or silo shape and height-to-diameter ratio for facilitating flow. It appears that a tall, slender structure is preferred to the short, squat one, with a height-to-diameter ratio in the range of 2.5:1 to 4:1 the most desirable. Good flowability is promoted

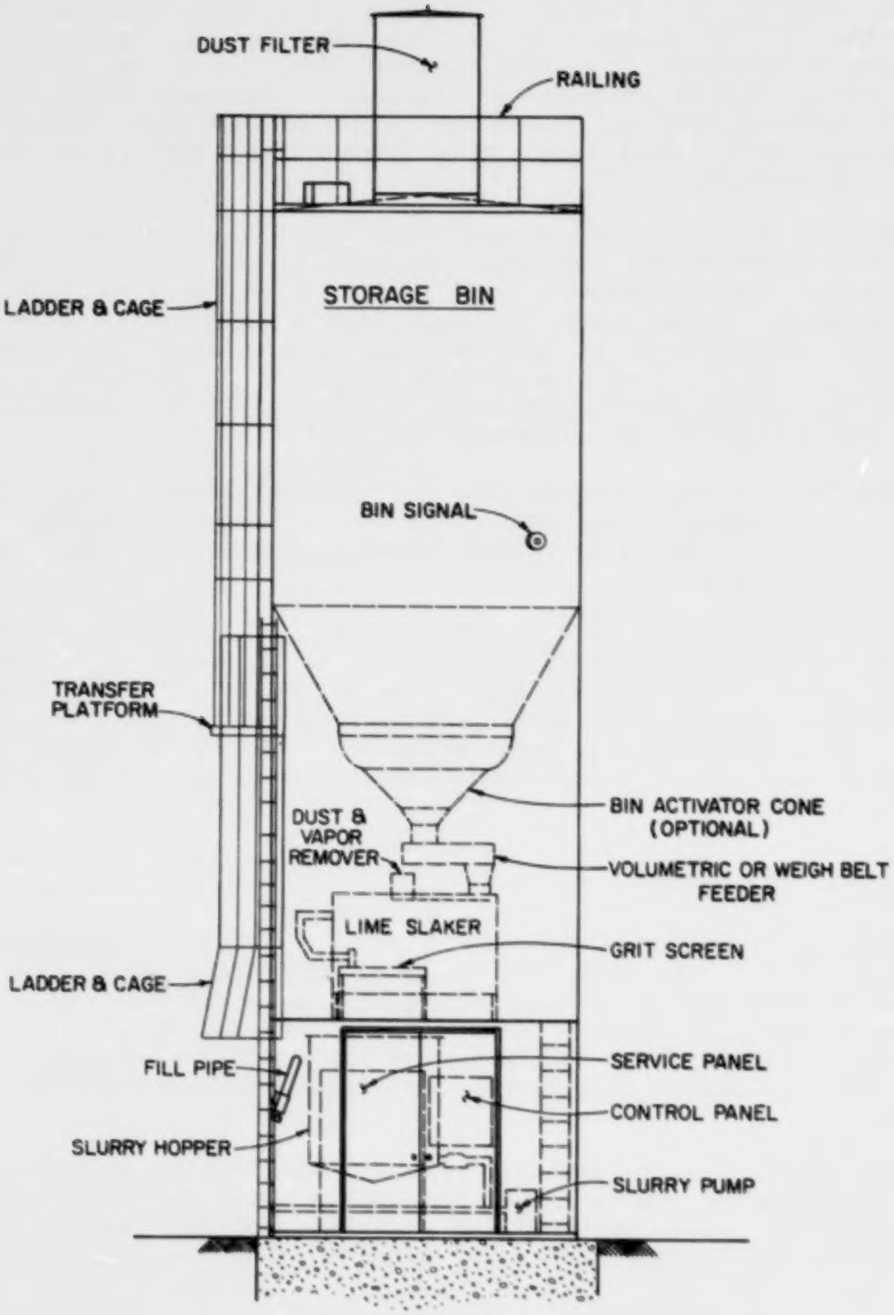

DUST FILTER
RAILING
STORAGE BIN
LADDER & CAGE
BIN SIGNAL
TRANSFER PLATFORM
BIN ACTIVATOR CONE (OPTIONAL)
DUST & VAPOR REMOVER
VOLUMETRIC OR WEIGH BELT FEEDER
LIME SLAKER
GRIT SCREEN
LADDER & CAGE
FILL PIPE
SERVICE PANEL
CONTROL PANEL
SLURRY HOPPER
SLURRY PUMP

Figure 3-1. Typical quicklime and slaker installation.

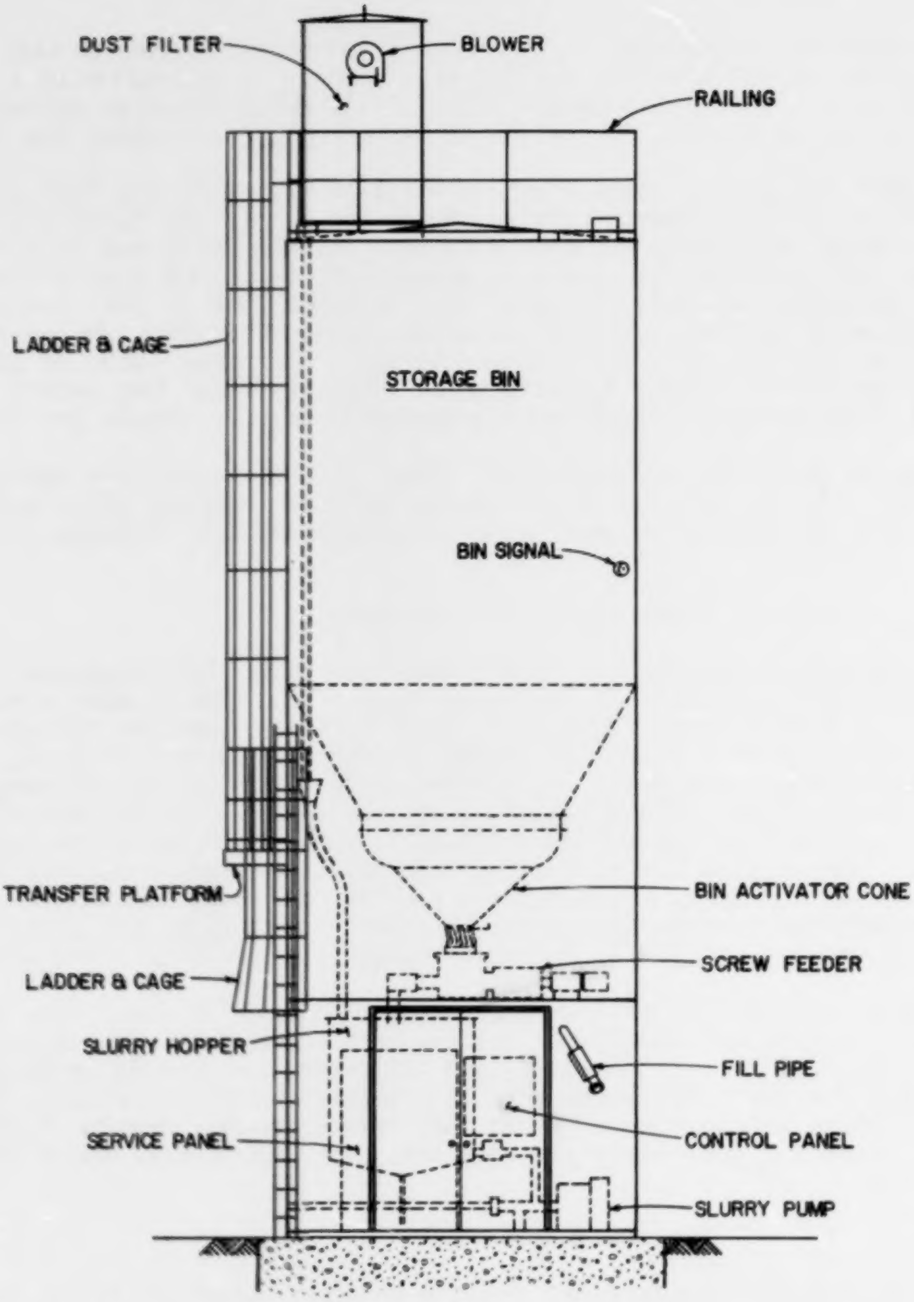

DUST FILTER
BLOWER
RAILING
LADDER & CAGE
STORAGE BIN
BIN SIGNAL
TRANSFER PLATFORM
BIN ACTIVATOR CONE
SCREW FEEDER
LADDER & CAGE
SLURRY HOPPER
FILL PIPE
SERVICE PANEL
CONTROL PANEL
SLURRY PUMP

Figure 3-2. Hydrated lime and slurry installation.

by having the discharge area as large as possible in relation to bin cross-sectional area.

With quicklime or hydrate, it is advisable for hopper bottoms to have a minimum slope of 60° from the horizontal. One way of accomplishing a steeper slope is to use an offset hopper (Figure 3-3), which serves to increase flowability by decreasing the weight of material going through the opening.

The need for such a steep slope is based on the relatively high angles of repose of 50°-55° (average) for quicklime and 15°-80° for hydrated lime. In the case of quicklime, the angle of repose is affected by the particle size, shape, and gradation (particularly percent of fines), and type of lime. The wide variation in angle of repose for hydrated lime is the result of the variation in particle size and gradation, moisture content, degree of aeration, or presence of electrostatic charges. The lower value of 15° would apply to highly aerated hydrate, which literally flows like water; in contrast, high moisture content and electrostatic charges produce the 80° value.

Generally, with the recommended 60° slope, quicklime will flow readily without the need for external aids, whereas hydrated lime and pulverized quicklime will require one or more means to promote uniform discharge from bins.

3.2.5.1 Vibrators or Bin Activators

The simplest device for improving flowability is an electromagnetic vibrator attached to the outside of the hopper face. This type is more suitable for quicklime than hydrate. Vibrators, however, can be used for hydrate if the unit is cycled to produce 1-2 seconds of vibration every 5-10 seconds. With quicklime, the vibrator can be operated continuously during discharge. For best results, the vibrator should be bolted directly to the conical hopper face, one-fourth or less of the distance from the discharge to the top of the cone. Vibrators should be operated only while the hopper is open to flow; this prevents packing. If noise is a problem, the electromechanical vibrator is recommended over the electromagnetic unit. Both types are available in a variety of sizes to fit the individual hopper size and metal plate thickness.

Air pads and jets are other means of inducing material flow. They are not, however, recommended for lime. They are best suited for highly dense materials such as iron ore. Vibratory bin activators are usually preferred over air pads for the lime systems used in treating acid mine drainage. The primary problem with air pads is that they reduce the bulk density of the lime, which interferes with feeder accuracy and, more importantly, causes flooding.

3.2.5.2 Live Bin Bottoms

Several companies manufacture "live bin bottoms," which are fitted to new or existing bin structures. These units are vibrated continuously during unloading by a gyrator exerting a horizontal force of up to 18,160 kg (40,000 lb). This promotes a steady flow of material with a uniform density to the feeder.

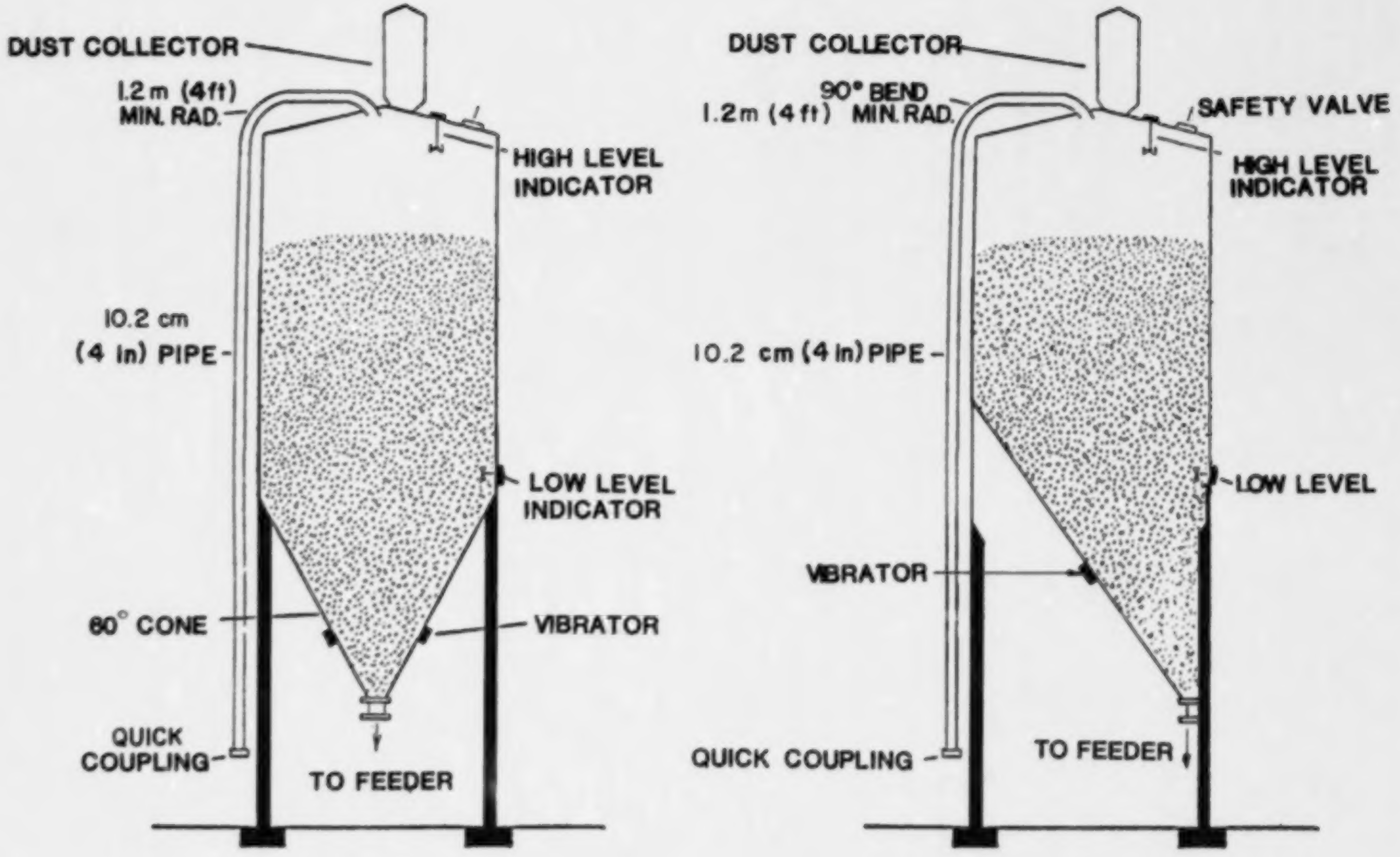

DUST COLLECTOR
1.2 m (4 ft)
MIN. RAD.
HIGH LEVEL INDICATOR
10.2 cm (4 in) PIPE
LOW LEVEL INDICATOR
60° CONE
VIBRATOR
QUICK COUPLING
TO FEEDER
DUST COLLECTOR
90° BEND
1.2 m (4 ft) MIN. RAD.
SAFETY VALVE
HIGH LEVEL INDICATOR
10.2 cm (4 in) PIPE
LOW LEVEL
VIBRATOR
QUICK COUPLING
TO FEEDER

Figure 3-3. Standard and offset hopper bottoms.

A device called the "bin activator" incorporates an internal dome-shaped baffle plate above the discharge, which exerts a 45,400-kg (100,000-lb) vertical thrust into the bin. Vibrations from the baffle plate penetrate the overlying material, forcing it to move freely to the periphery of the plate, then down to the discharge. Bridging and ratholing are virtually eliminated. The bin activator is available in 0.6- to 3.0-m (2- to 10-ft) diameter sizes and is usually sized one-half the total diameter of a circular silo.

3.2.6 Lime Feeders

Existing AMD treatment processes utilize dry lime in a liquid suspension or slurry before introducing it into the raw water. Today, however, designers are beginning to feed dry hydrated lime directly to the acid water. This eliminates a slurry feed system. Dry lime feed has been used when the drainage streams are small and mildly acidic. These conditions require little lime (less than 0.1 kg/1,000 l), and past designers have installed such systems because of economics. The primary reason is that dry hydrated lime can be employed this way with an acceptable degree of efficiency.

For mine drainage requiring larger amounts of lime per unit flow, 0.36-0.48 kg/1,000 l (3-4 lb/1,000 gal), special design considerations must be provided. An example would be a larger aeration or flash mixer to insure complete mix and utilization of the lime. This method of lime feeding is further explained in Chapter 12.

The lime feeder selection for a treatment plant depends largely on the type and size of lime specified and daily lime requirements (generally figured in kg/hr (lb/hr) or m^3/hr (ft^3/hr)). Regardless of whether a volumetric or gravimetric feeder is used, one should be selected that provides sufficient flexibility, protection against exposure to lime dust, low maintenance, few moving parts, and ease of cleaning. Some feeders are designed especially for granular materials such as pebble lime, others for powdery materials such as hydrate.

A machine that requires frequent maintenance is undesirable in an efficient, continuous, automated process. Easily cloggable mechanisms, such as plungers and plates using small orifices or clearances, have been replaced by the oscillating hopper, belt conveyor, and vibrating feeder. These usually have large clearances at all points, so larger grains and small pieces of paper or wire pass the machines. Many operators take the precaution of placing a screen across the hopper opening to remove paper and wire during loading. This reduces the chances of blockage, and helps to maintain a dry hydrated lime of constant density within the feeder hopper.

All feeders should be capable of confining dust for the sake of cleanliness, health, economy, and compliance with increasingly stringent air pollution and OSHA regulations. Any feeding mechanism that cannot be easily and conveniently enclosed may create a dust problem. Fortunately, most feeds are designed with this in mind. Consideration also should be given to necessary shutdowns for repair, maintenance, and adjustments to the feed. Moving parts of mechanical equipment must be lubricated regularly, and electrical equipment must be replaced from time to time.

3.2.6.1 Dry Feeders

The efficiency of an acid mine drainage treatment plant depends on the speed and accuracy with which the lime is handled and fed to the process. To maintain the proper function of the neutralization process, it is essential that an uninterrupted flow of lime be maintained upon demand. The importance of operating lime feeders properly cannot be overly stressed.

The dry lime feeder consists of two parts:

1. a feeder hopper, usually with a throat at the bottom through which the lime falls by gravity;

2. a feeding element (i.e., screw) that can be adjusted to give different rates of material. The feed may be by volume or by weight; i.e., volumetric or gravimetric. A volumetric feeder will deliver a constant volume, but the weight may vary.

With dry feeders, the decision to use volumetric or gravimetric will depend upon desired accuracy and economics. The accuracy of gravimetric feeders is not needed for treating acid mine drainage. Volumetric feeders, which may have an error of 7%-15% by weight, are more than sufficient. This error is dependent upon the flowability and uniform density of the lime. In most lime feeding systems, the lime requirements are controlled more by the method of feeding the solution than by the accuracy of the feeder itself. This is probably the reason for the popularity of volumetric feeders in AMD plants.

The type of lime used in the feeder is an important factor. A feeder that may introduce pebble lime with accuracy may not do so with hydrated lime: hydrated lime is more difficult to feed.

a. Feeder Hopper

A dry feeder will operate only as well as the hopper that charges material to it. Therefore, it is important that the material flows freely and uniformly from the hopper. Several techniques have already been discussed under bulk silo storage. The following discussion presents additional information useful to the designer and treatment plant operator.

Most feeders are equipped with a standard hopper, the size of which depends on the capacity of the feeder. Feeder hoppers are universally designed so that the slope angle of the hopper bottom is greater than the angle of repose of the lime to be fed from it. The slope angle, although varying slightly from model to model, is usually about 60°. Hoppers for feeders are usually conical or rectangular (see Figure 3-4 A). No matter what shape, hydrated lime may arch or bridge.

When the material arches in an unagitated hopper, the operator is powerless to do anything about it except resort to such primitive methods as pounding the side of the hopper, or going on top of the hopper and agitating the mass with a stick or paddle. None are effective or acceptable methods. Measures should be taken to prevent the arching from ever occurring, although no method is absolutely foolproof.

b. Volumetric Feeders

As mentioned earlier, volumetric feeders supply a constant, preset delivery of material by volume and do not recognize changes in material density. Consequently, this type of feeder must be calibrated by trial and error at the outset, then readjusted periodically if the lime changes in density.

There are more than a dozen types of volumetric feeders on the market. Only five, however, are suited to feeding lime in the quantities required for treatment of acid mine drainage. Of these, the screw-type conveyors are most commonly used for pebble lime, and the rotary paddle or star-type feeder for hydrated lime. The following sections discuss these feeders and present details on specific models.

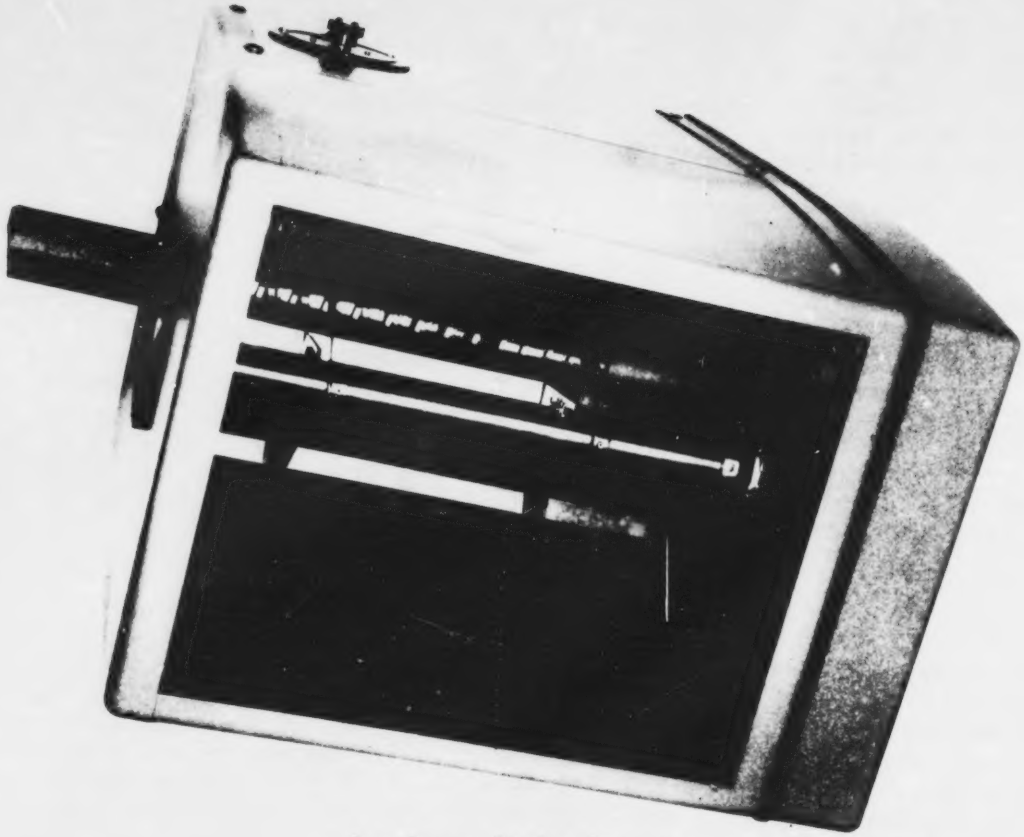
c. Screw Feeder

The screw conveyor feeder (Figure 3-4 B) delivers a constant stream of material from the hopper. For solid screw-type conveyors, usually constant-pitch flights are employed, but tapered or variable-pitch flights can also be used. At the same time, the load on the screw should not be so excessive as to prevent turning. The capacity of a screw feeder can readily be changed by varying the speed of the shaft. It is highly recommended that a variable-speed screw feeder be installed. This provides the operator with better control and flexibility for seasonal variations of lime demand.

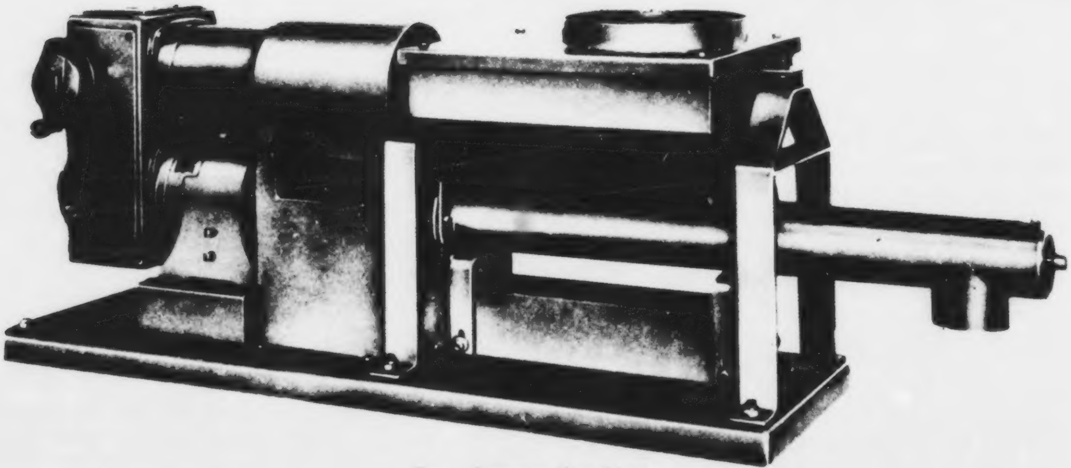
A recent development in volumetric screw feeding is the vibrating screw. The feeder assembly is subjected to continuous, controlled gyratory vibration, which insures that each flight of the screw conveyor will be filled to the maximum and that each will be completely emptied at the discharge tube. There are two models: (1) a heavy-duty feeder, with feed rates from as low as 0.0078 m^3/hr (0.28 ft^3/hr) for a 2.5-cm (1-in) screw size to as high as 16.8 m^3/hr (600 ft^3/hr) for the 15.2-cm (6-in) screw; and (2) the live bin screw feeder, with rates varying from 1.0×10^{-4} m^3/hr (0.0037 ft^3/hr) for a 0.64-cm (0.25-in) screw to 5.6 m^3/hr (200 ft^3/hr) for a 10.2-cm (4-in) screw. With the latter unit, the entire bin conveyor assembly is subjected to gyratory vibration to insure undiminished flow from hopper to discharge. The manufacturer claims acceptable feeder accuracy.

d. Oscillating Hopper

The oscillating hopper feeder (Figure 3-5) consists of an oscillating hopper that swivels on the end of the main hopper. The material completely fills both hoppers and rests on the tray beneath. As the oscillating hopper moves



A. Screw feeder hopper.



B. Screw feeder.

Figure 3-4. Lime feeders.

back and forth, the scraper, which rests on the fixed tray below, is moved first to the left and then to the right. As it moves, it pushes a ribbonlike layer of material off the tray. The capacity is fixed by the length of the stroke, which may be varied by means of a micrometer screw. Further adjustment is possible by changing the clearance between the hopper and the tray, which can be raised or lowered. This type of feeder is one of the most widely used in small installations.

e. Belt Feeder

With a belt feeder, the material enters the feed section from an overhead hopper, falls on the feed belt, and passes beneath a vertical gate. For a given belt speed, the position of the gate determines the volume of material passing through the feeder (see Figure 3-6). With one particular feeder, the gate is manually positioned by a cam, and the setting indicated by dial and pointer. A 22.8-cm (9-in) belt feeder is available with a 16.8-m^3/hr (600-ft^3/hr) capacity. The feed rate is adjustable over a basic range of 10 to 1 (100 to 1 optional). This particular belt feeder can accommodate pebble lime of maximum 3.8-cm (1.5-in) size.

f. Rotary Paddle

The rotary paddle feeder is particularly effective for fine materials that tend to flood, such as hydrated lime. The paddle or vane is located beneath the hopper discharge, with the feed varied by means of a sliding gate and/or variable-speed drive for the paddle shaft. The pocket feeder, also called the star or revolving-door feeder, where the paddle is tightly housed, permits delivery against vacuum or pressure. One such star feeder made especially for hydrated lime and soda ash incorporates tight-fitting neoprene blades that control floodable materials. Star feeders work best when coupled with a live bottom bin.

g. Vibrating Feeder

The vibrating feeder obtains motion by means of an electromagnet anchored to the feeding trough, which in turn is mounted on flexible leaf springs. The magnet, energized by a pulsating current, pulls the trough sharply down and back, then the leaf springs return it up and forward to its original position. This action is repeated 3,600 times/min (when operating on 60-cycle AC), producing a smooth, steady flow of material.

3.2.6.2 Prevention of Feeder Flooding

Feeder flooding is closely related to arching and is caused by a sudden breaking of an arch or otherwise clogged state. The theories on why this happens vary; however, one theory states that the material entrains air, becomes fluidized, and floods (literally gushes), thereby inundating the feeder. This tendency is much more pronounced in light fluffy materials,

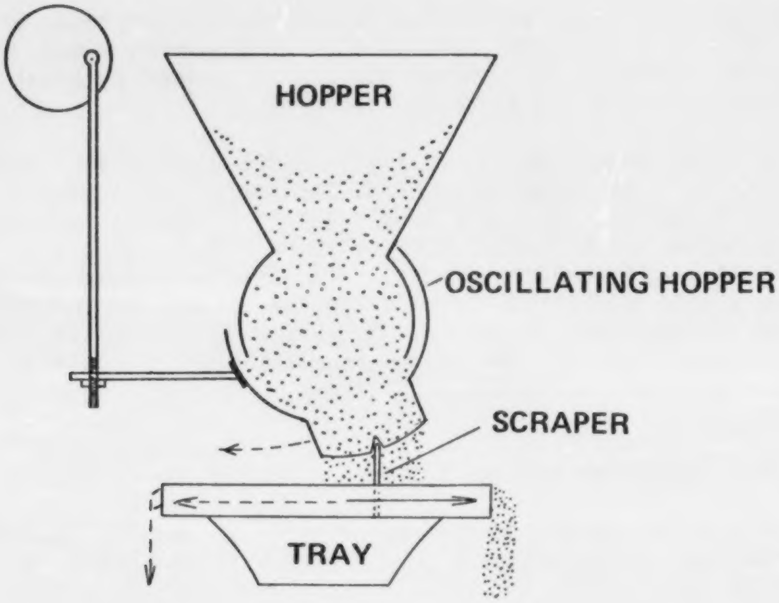

HOPPER
OSCILLATING HOPPER
SCRAPER
TRAY

Figure 3-5. Oscillating hopper feeder.

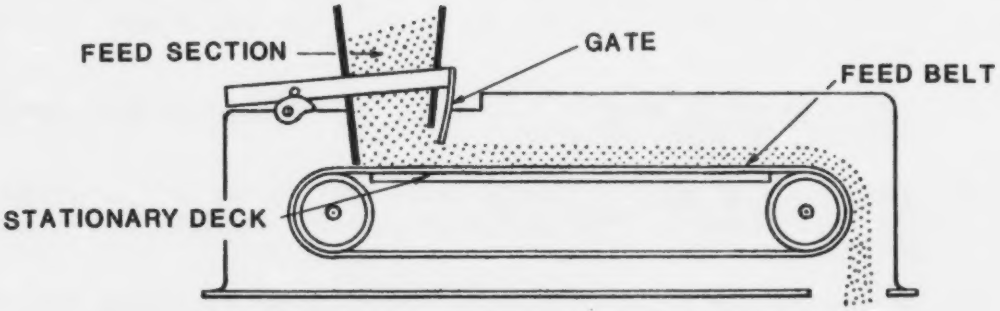

FEED SECTION
GATE
FEED BELT
STATIONARY DECK

Figure 3-6. Belt feeder.

such as hydrated lime, which flow freely and are sometimes difficult to control. This fact seems to add credence to the entrainment theory; for this reason, the practice of blowing air into a hopper to break an arch can be considered an invitation to feeder flooding.

Flooding often occurs when new lime is added to an almost empty hopper. The effect of the lime dropping through the hopper and probably entraining air may cause flooding. The silo hopper should always be equipped with either a bin gate valve or a feeder flood shutoff. Most vendors will supply flood shutoffs at a slight extra charge. The primary purpose of the slide valve, which is always positioned between the feeder and bin activator, is to prevent the uncontrolled dispersion of lime when filling an empty silo. Many times, several feet of lime empty onto the slurry or slaker room floor because the designer overlooked this item.

3.2.7 Quicklime Slaking Systems

The term slaking refers to the combination of varying proportions of water and quicklime, which yields a milk of lime (lime slurry) or a viscous lime paste with some degree of consistency. For maximum efficiency when using quicklime, it is desirable to slake the lime at or near optimum conditions. Since many limes have different slaking characteristics, the optimum conditions are usually determined by trial and error. Most lime suppliers can simplify this determination by providing data or recommendations on the slaking behavior of their quicklime. The way the quicklime is slaked can mean the difference between an efficient and a wasteful operation.

Briefly, the variables exerting a profound effect on slaked hydrate quality are:

1. Reactivity of the quicklime - The reactivity of the quicklime depends upon whether the quicklime is hard-, soft-, or medium-burned.

2. Particle size and gradation of quicklime - No matter what grade of quicklime is used (lump, pebble, ground, pulverized, or run-of-the-mill), the finer sizes slake most rapidly because of more available wetting area.

3. Optimum amount of water - Too much or too little water will slow the reaction and reduce efficiency.

4. Temperature of water - Slaking water that is too cold or possibly too hot (steam) for the particular slaking conditions also slows the reaction.

5. Distribution of water - An even flow of water introduced into the slaking chamber is highly recommended.

6. Agitation - Too vigorous or insufficient agitation of quicklime and water will result in a poor and hazardous operation.

The heat of the slaking reaction is important (average 88°C (191°F)), and the reaction can be artificially accelerated by using hot water. By such measures, it may be possible to increase the slaking rate of a medium reactive lime to the approximate behavior of a high reactive lime. When applying this method to a reactive lime, it is possible to obtain extremely rapid, almost instantaneous slaking, so the lime and water literally explode on contact. Striving for such explosive slaking as this, however, is inadvisable. A complete slaking time of 5-10 minutes at a rapid and uniform rate is considerably more desirable.

Two extreme conditions should be avoided. If excessive quantities of slaking water are used, particularly cold water, "drowning" occurs. The surface of the quicklime particle hydrates quickly, but the mass of hydrate formed impedes the penetration of the water into the center of the particle, delaying explosion of the particle into microparticles. The rise in temperature is stifled and slaking delayed, resulting in coarser hydrate particles and incomplete hydration. The other extreme is adding insufficient water to the lime, causing the hydrate to be "burned" because of excessive temperatures, 121°-260°C (250°-500°F), instead of the desired temperature just below boiling, 88°C (191°F). Much of the hydration water is lost as steam; thus, a considerable number of nonhydrated particles remain. Also, the heat can be so intense that paint on the equipment can blister or ignite, and lime parti cles initially hydrated can be dehydrated.

3.2.7.1 Continuous Slaking

Because of the obvious economy and efficiency in slaking under optimum or near-optimum conditions, manually operated batch slaking has been largely replaced by continuous slakers. There are two basic types of slakers: (1) the detention type that produces a lime slurry or creamy suspension; and (2) paste slakers that produce only a paste or putty. Both types are generally equipped with dilution tanks so any desired concentration of lime slurry can be made.

There is some variance in the proportion of water and lime used, depending on the characteristics of the lime and the type of slaker. With detention slakers, the lime-to-water (weight) ratio averages 3:1 to 4:1 for high calcium quicklimes and 2:1 for paste slakers. Most detention slakers will handle a quicklime of 5 cm (2 in) top size down to pulverized forms, but paste slakers require nothing larger than 19 mm (0.75 in). The trend today is toward the use of smaller sizes of fairly restricted gradations of 19 mm x 9.5 mm (0.75 in x 0.38 in) or smaller granular and pebble sizes. The principle of either type of slaker is the same; i.e., to interact quicklime and water with sufficient contact time to achieve complete hydration while continuously discharging a slaked lime from the vessel.

The major difference between these two types is that the paste slaker operates more easily at higher temperatures, 88°-99°C (190°-210°F), than the detention slaker, because more heat is generated from the lower water-to-lime ratio. To achieve the same desired temperature in a detention slaker, it is often necessary to augment the natural heat of hydration by one or more of the following methods: (1) using hot water for slaking; (2) using a heat exchanger in a dual tank system to capture some of the heat of hydration for use in the first slaking tank; (3) using heavy insulation around the slaking compartment to reduce heat losses; (4) using a longer retention time to complete hydration. Paste slakers require only 5-10 minutes for complete hydration as opposed to 20-30 minutes for the detention slaker.

Proponents of the detention-type slaker claim that this unit can slake poorer quality, slower slaking limes more efficiently, offering more flexibility in accommodating the whole spectrum of limes. Another possible advantage for the detention type is the prospect of longer life and less downtime and maintenance.

Two popular continuous slaker systems are illustrated in Figures 3-7 and 3-8. These units provide two separate compartments: one for initially slaking to 35%-40% solids (putty consistency), and the other for dilution that produces a workable slurry or milk-of-lime solution (5%-10%). Each of these compartments is equipped with rotating paddles for agitation and complete mixing. Also, the basic quicklime slaking unit provides a shaker or oscillating screen for slurry degritting, a hood for vapor and dust removal, and the necessary pumps and piping for transporting the slurry to the point of application.

Another widely used slaker is shown in Figure 3-9. This continuous detention-type slaker provides dual propeller-type mixers, one for each compartment, and a thermostatically controlled slaking temperature. This unit produces a fairly consistent slurry and provides flexibility for an operator's needs in AMD treatment. As standard procedure, a thick pastelike slurry is made in the first compartment, then diluted and retained for complete hydration in the second. The desired slurry (solids percent) is controlled by a preset water-to-quicklime ratio.

The last compartment has a separator trough equipped with a screw raking unit on an incline for the removal and discharge of grit. This process enables the grit to be loaded directly into a disposal or transporting container.

Most slakers are available in three and four compartments with a wide range of capacities, 3.2-45.4 kkg/24 hr (3.5-50 tons/24 hr). Also, they are equipped with heat exchangers for heat conservation and temperature control.

As mentioned previously, a slaking temperature ranging from 82°-88°C (180°-190°F) is important, and it can be maintained automatically by a thermostatically controlled water valve. Generally, the valves respond to a 3°C (5°F) temperature change by adding varying amounts of water. Often, modern slakers are insulated to retain the heat of hydration, which enhances operation during seasonal weather variations.

When quicklime is selected as the neutralizing agent, grit removal must be undertaken to preserve pump and equipment (tanks, pipes, valves) life. Even the highest quality quicklimes contain 1.5%-3.0% grit. The grit is composed of silica, alumina, carbonate core, and insoluble calcium compounds. Based on a lime usage of 18.2 kkg/d (20 tons/d), approximately 273-545 kg (600-

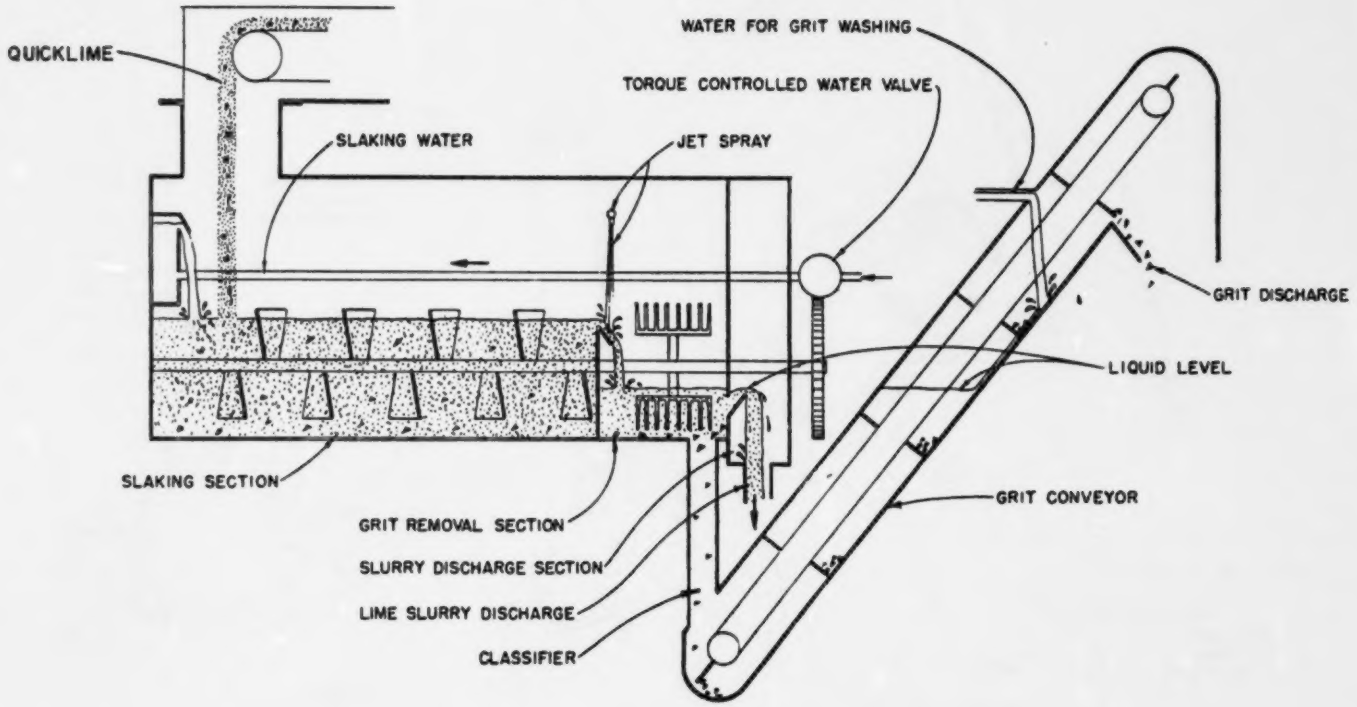

QUICKLIME
SLAKING WATER
JET SPRAY
WATER FOR GRIT WASHING
TORQUE CONTROLLED WATER VALVE
GRIT DISCHARGE
LIQUID LEVEL
GRIT CONVEYOR
SLAKING SECTION
GRIT REMOVAL SECTION
SLURRY DISCHARGE SECTION
LIME SLURRY DISCHARGE
CLASSIFIER

Figure 3-7. Paste slaker with classifier for grit removal.

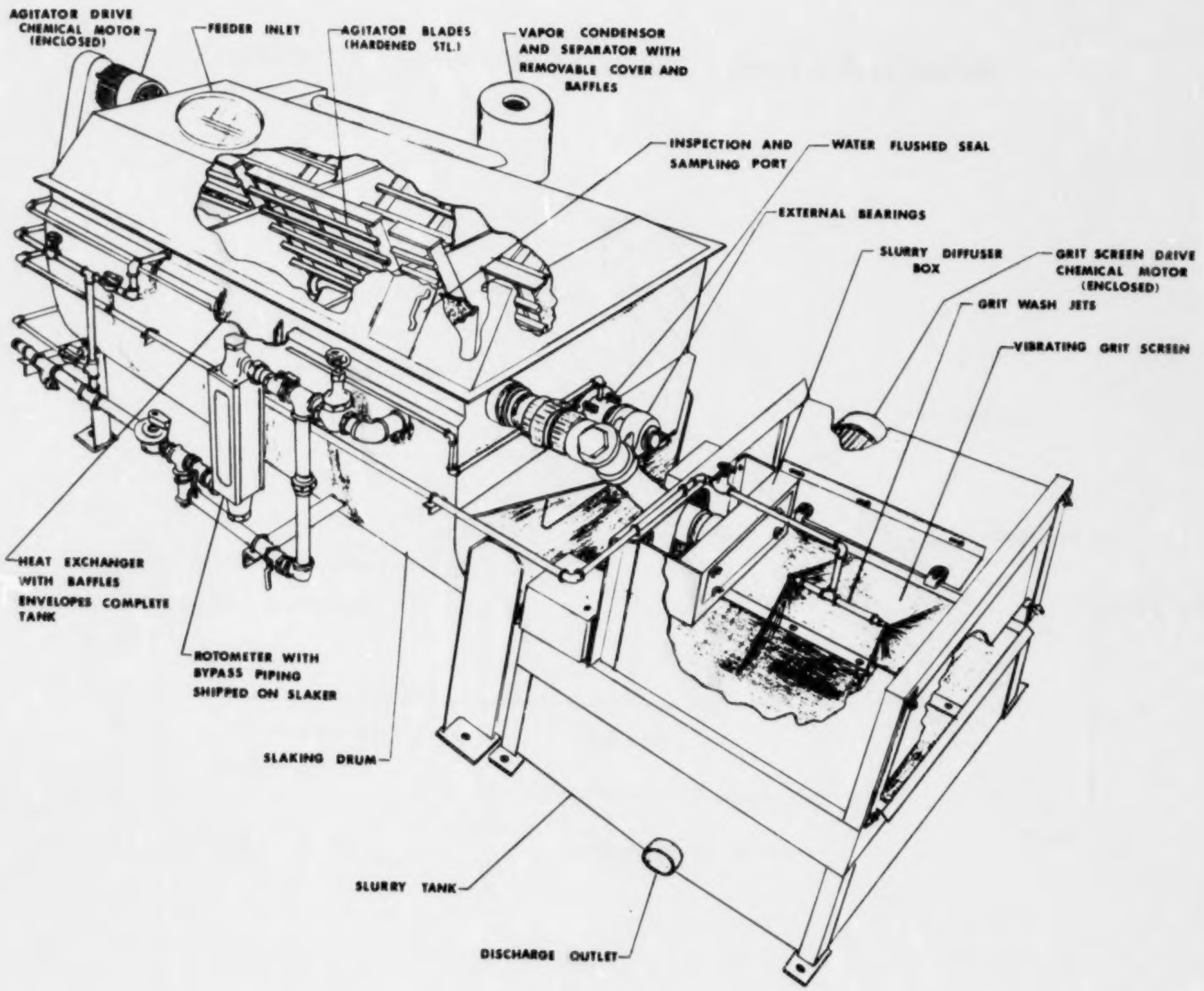

AGITATOR DRIVE CHEMICAL MOTOR (ENCLOSED)
FEEDER INLET
AGITATOR BLADES (HARDENED STL.)
VAPOR CONDENSOR AND SEPARATOR WITH REMOVABLE COVER AND BAFFLES
INSPECTION AND SAMPLING PORT
WATER FLUSHED SEAL
EXTERNAL BEARINGS
SLURRY DIFFUSER BOX
GRIT SCREEN DRIVE CHEMICAL MOTOR (ENCLOSED)
GRIT WASH JETS
VIBRATING GRIT SCREEN
HEAT EXCHANGER WITH BAFFLES ENVELOPES COMPLETE TANK
ROTOMETER WITH BYPASS PIPING SHIPPED ON SLAKER
SLAKING DRUM
SLURRY TANK
DISCHARGE OUTLET

Figure 3-8. Paste slaker with vibrating screen for grit removal.

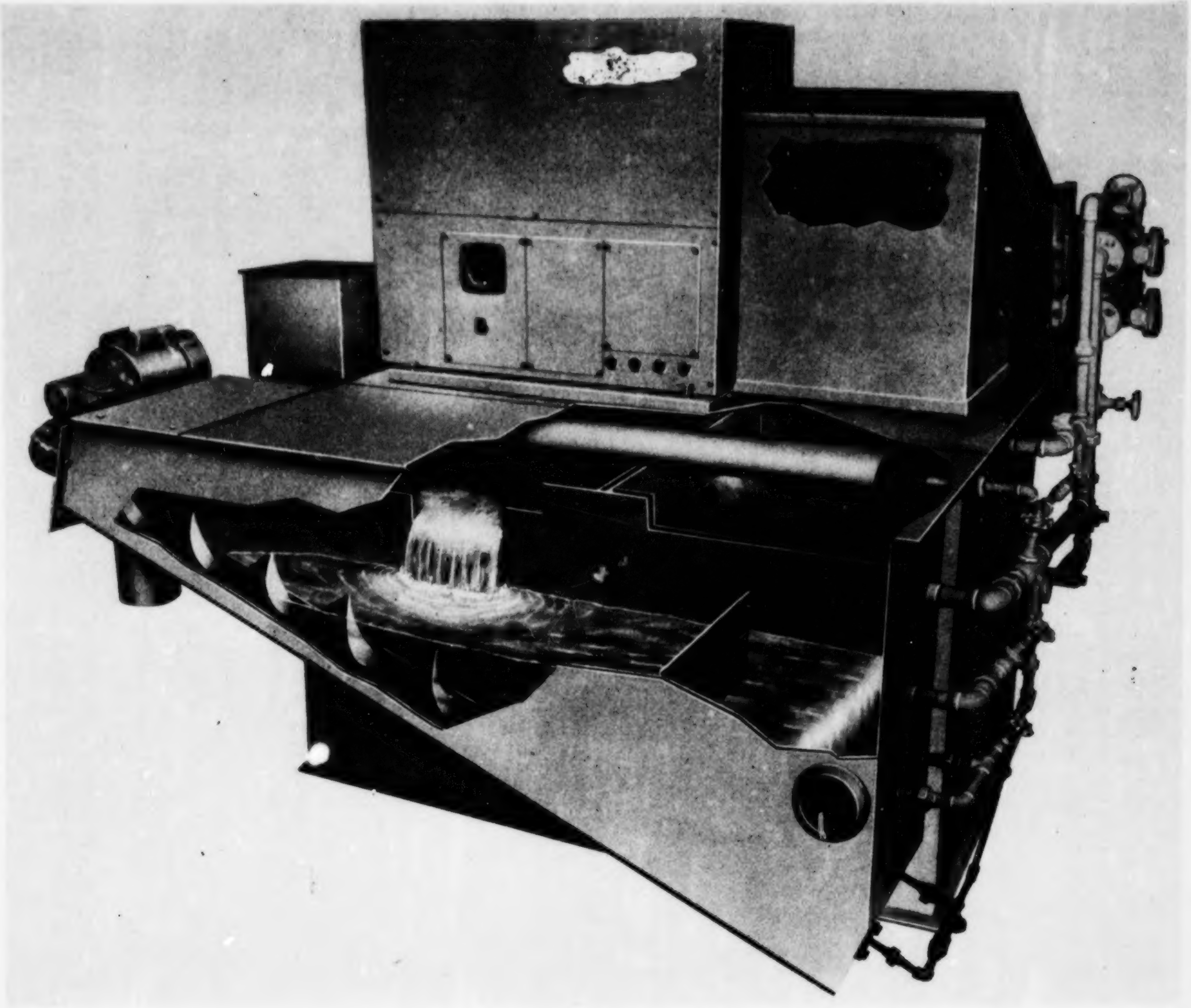


Figure 3-9. Detention slaker.

1,200 lb) of grit will require daily disposal.

The designer should be aware of the utilities required for a lime slaker. Sometimes three-phase power or pneumatic air is needed. Slaking water is most important with a clean supply always preferred. Past practice, however, has been to use plant effluent, although in some cases the raw drainage is used. This practice must be exercised with caution and the dissolved ion chemistry carefully reviewed to prevent gypsum formation or retarding reaction.

Research to show the importance of water quality in efficient lime slaking proved that water of or near potable quality is most desirable. In particular, it was demonstrated that wastewater or recycle-process water containing sulfites and sulfates retarded the slaking process. Not only was more time needed to complete slaking, but the quality of the resulting lime slurry was impaired. The lime hydrate particle size became much larger and the surface area smaller, which, in turn, retarded the neutralization reaction with acids. In fact, some of the lime did not hydrate and was wasted. The main explanation is that lime precipitates the sulfite and sulfate ions that coat the unreacted calcium oxide, preventing complete water penetration into the particles.

It was also discovered that the wastewater or recycle-process water can be used after slaking to dilute the thick lime slurry to the desired consistency. The effect of the sulfite and sulfate ions on the quality of the diluted lime slurry was negligible. The chloride ion in reasonable amounts (500 mg/l) did not appear to exert any deleterious effect on slaking, but higher concentrations retarded the reaction.

3.2.7.2 Degritting of Quicklime Slurries

Even the highest quality quicklimes have a grit content ranging between 1.5% and 3.0% of the weight. Included in the grit, along with the carbonate core, are insoluble silicates, aluminates, sulfates, ferrites, and all impurities in the limestone before the lime was calcined. When the grit is ejected from the slaker, it resembles a mass of wet sand with particles ranging from 6.4 mm (0.25 in) in diameter to #100 mesh.

Degritting improves lime quality and reduces abrasion and wear on equipment. In extreme cases, cast iron centrifugal pumps have been worn out within a month when pumping a slurry that has not been degritted. With efficient degritting, the same equipment can operate for years without maintenance.

Degritting is performed in the dilution or stabilization tank adjacent to the slaking chamber. As the slurry or paste passes over a weir into the dilution chamber, it is dispersed and diluted by water sprays. The heavier grit particles settle rapidly to the bottom and are removed automatically by rakes. In some slakers, there is a classifier in the bottom of the dilution tank where the grit is washed and the washwater recovered.

The washed grit is then disposed of manually or automatically. Slakers with a capacity of 227 kg/hr (500 lb/hr) or more usually employ automatic eductors. Meanwhile, enough turbulence is maintained so that nearly all of the slaked lime in the diluted slurry remains in suspension and is piped to storage.

3.2.7.3 Quicklime Slurry Concentrations

Designer preference as to the concentration of lime solids in the milk-of-lime slurry used in their process varies from 5% to 20%. Most designers will dilute the slurry to at least a 10% lime solids concentration before it leaves the slaker. There is no fixed rule; however, it is recommended that slurry feed systems be designed at a 10% lime-to-water weight ratio because a more concentrated slurry causes additional maintenance and operational problems. In any event, the slurry feed rate should never exceed more than 10% of the drainage flow. The concentration of lime can be checked for specific gravity with a hydrometer and by using the data in Table 3-1.

The use of automatic pH control systems for the feeding of lime solutions into the flash mix tank has become common in recent years. Therefore, the strength of the slurry is not as important as the volume of slurry being fed. Regardless of the approach used in controlling the lime feed, it is necessary to maintain the lime in suspension and prevent settling to maintain a uniform feed.

Also, it is important to recognize that percent lime solids is not equivalent to percent calcium oxide. The percent lime solids refers strictly to the hydroxide ($Ca(OH)_2$) and not the oxide (CaO). Table 3-1 shows the equivalent of $Ca(OH)_2$ as CaO. To determine what concentration of lime solids exists, the following simple equation is used:

$$\% \text{ Solution} = \frac{\text{kg of } Ca(OH)_2}{\text{kg of } Ca(OH)_2 + (1.0 \text{ kg/l} \times \text{No. of l of water})} \times 100 \qquad (4)$$

or

$$\% \text{ Solution} = \frac{\text{lb of } Ca(OH)_2}{\text{lb of } Ca(OH)_2 + (8.345 \text{ lb/gal} \times \text{No. of gal of water})} \times 100$$

3.2.8 Slurried Lime Transfer

Usually the diluted paste or slurry ready for use must be transferred a short distance to the mixing or neutralization tank. It is here that scaling can become a serious problem. Lime is slightly soluble at best; a saturated solution is 1.7 g/l (0.14 lb/gal) at 0°-10°C (32°-50°F). As the temperature rises, the solubility of lime decreases until at 90°-100°C (194°-212°F) it is only 0.55 g/l (0.005 lb/gal). Thus, it is economically essential to convey lime in a much more concentrated form such as a suspension. Because lime slurry has a pH greater than 12.0, the water that carries it undergoes a softening action and precipitates fresh calcium carbonate as a dense, hard

TABLE 3-1

PROPERTIES OF LIME SLURRIES

Approx. % Solids $Ca(OH)_2$ (by wt)	Specific Gravity @ 15°C	Pounds per gallon	Grams CaO per liter	Grams $Ca(OH)_2$ per liter	Pounds CaO per gallon	Pounds $Ca(OH)_2$ per gallon	Pounds Slurry per cubic foot	Weight Ratio Water to CaO	Weight Ratio Water to $Ca(OH)_2$
1.5	1.010	8.41	11.7	15.46	0.097	0.128	62.94	85.70:1	64.70:1
3.2	1.020	8.50	24.4	32.24	0.203	0.268	63.62	40.81:1	30.72:1
4.8	1.030	8.58	37.1	49.02	0.309	0.408	64.22	26.77:1	20.03:1
6.3	1.040	8.66	49.8	65.81	0.415	0.548	64.82	19.87:1	14.80:1
7.8	1.050	8.75	62.5	82.59	0.520	0.686	65.49	15.83:1	11.76:1
9.4	1.060	8.83	75.2	99.37	0.626	0.826	66.09	13.11:1	9.69:1
10.8	1.070	8.91	87.9	116.15	0.732	0.966	66.69	11.17:1	8.22:1
12.2	1.080	8.99	100.0	132.14	0.833	1.10	67.29	9.79:1	7.17:1
13.7	1.090	9.08	113	149.32	0.941	1.24	67.96	8.65:1	6.32:1
15.2	1.100	9.16	126	166.50	1.05	1.39	68.56	7.72:1	5.59:1
16.4	1.110	9.25	138	182.35	1.15	1.52	69.23	7.04:1	5.09:1
18.0	1.120	9.33	152	200.85	1.27	1.68	69.83	6.35:1	4.55:1
19.2	1.130	9.41	164	216.71	1.37	1.81	70.43	5.87:1	4.20:1
20.4	1.140	9.50	177	233.89	1.47	1.94	71.10	5.46:1	3.90:1
21.8	1.150	9.58	190	251.07	1.58	2.09	71.73	5.06:1	3.58:1
23.1	1.160	9.66	203	268.24	1.69	2.23	72.35	4.72:1	3.33:1
24.4	1.170	9.75	216	285.42	1.80	2.38	72.97	4.42:1	3.10:1
25.6	1.180	9.85	229	302.60	1.91	2.52	73.73	4.15:1	2.91:1
26.9	1.190	9.91	242	319.78	2.02	2.67	74.17	3.91:1	2.71:1
28.0	1.200	10.00	255	336.96	2.12	2.80	74.85	3.72:1	2.57:1
29.2	1.210	10.08	268	354.14	2.23	2.94	75.44	3.52:1	2.43:1
30.4	1.220	10.16	281	371.31	2.34	3.09	76.04	3.34:1	2.29:1
31.5	1.230	10.24	294	388.49	2.45	3.23	76.64	3.18:1	2.17:1

(continued)

TABLE 3-1 (continued)

Approx. % Solids $Ca(OH)_2$ (by wt)	Specific Gravity @ 15°C	Pounds per gallon	Grams CaO per liter	Grams $Ca(OH)_2$ per liter	Pounds CaO per gallon	Pounds $Ca(OH)_2$ per gallon	Pounds Slurry per cubic foot	Weight Ratio Water to CaO	Weight Ratio Water to $Ca(OH)_2$
32.7	1.240	10.33	307	405.67	2.56	3.38	77.33	3.04:1	2.20:1
33.8	1.250	10.41	321	424.17	2.67	3.52	77.92	2.90:1	1.96:1
35.4	1.260	10.49	331	437.38	2.81	3.71	78.52	2.73:1	1.83:1
36.4	1.270	10.58	343	453.24	2.92	3.85	79.19	2.62:1	1.75:1
37.5	1.280	10.66	356	470.42	3.03	4.00	79.79	2.52:1	1.67:1
38.5	1.290	10.74	370	488.92	3.14	4.14	80.38	2.42:1	1.59:1
39.6	1.300	10.83	382	504.77	3.25	4.29	81.06	2.33:1	1.52:1
40.8	1.310	10.91	396	523.27	3.37	4.45	81.66	2.24:1	1.45:1
41.7	1.320	11.00	410	541.77	3.48	4.59	82.33	2.16:1	1.40:1
42.7	1.330	11.08	422	557.63	3.58	4.73	82.93	2.09:1	1.34:1
43.7	1.340	11.16	435	574.81	3.70	4.88	83.53	2.02:1	1.29:1
44.7	1.350	11.25	448	591.99	3.81	5.03	84.20	1.95:1	1.24:1
45.7	1.360	11.33	460	607.84	3.92	5.18	84.80	1.89:1	1.19:1
46.6	1.370	11.41	472	623.70	4.03	5.32	85.40	1.83:1	1.14:1
47.7	1.380	11.50	484	639.56	4.15	5.48	86.07	1.77:1	1.10:1
48.4	1.390	11.58	496	655.41	4.25	5.61	86.67	1.72:1	1.06:1
49.5	1.400	11.66	510	673.91	4.37	5.77	87.27	1.67:1	1.02:1
50.6	1.410	11.75	524	692.41	4.50	5.94	87.95	1.61:1	0.98:1
51.5	1.420	11.83	538	710.91	4.61	6.09	88.54	1.57:1	0.94:1
52.2	1.430	11.91	550	726.77	4.71	6.22	89.14	1.53:1	0.91:1
52.8	1.440	12.00	562	742.63	4.82	6.34	89.82	1.49:1	0.89:1
53.9	1.450	12.08	575	759.81	4.93	6.51	90.42	1.45:1	0.86:1

scale. If unattended, scale will build up and clog pipes. Most often, scale accumulates at the termination of the line to the treatment tank.

There is no foolproof solution to the problem; however, there are a few corrective measures that will minimize the problem and often prevent excessive maintenance. Those designing new plants should give serious thought to this problem because it may be possible to preclude scaling by appropriate design.

Scaling can also be prevented or minimized chemically. Sodium hexametaphosphate can be used to soften the slaking or dilution water so that the calcium carbonate that precipitates does not accumulate or scale in the pipe.

The best method is to locate the feeder so that the slurry flows by gravity directly into the solution or mixing tank. While this might not always be possible, designers should strive to accomplish this in the arrangement of the lime-feeding equipment. Scaling at the end of the slurry pipe can be avoided by discharging slurry through an air gap into an open solution tank. The use of heavy-duty, flexible rubber hoses or plastic pipes should be considered instead of metal pipes. These can be flexed or rapped to loosen any scale buildup. Using open troughs to convey the lime suspension makes scale removal simpler.

Much has been written about the components in a slurry feed system such as piping, valves, pumps, and configurations. The designer, however, should be aware of "practical engineering." That is, design with materials suitable for the application and with operator/maintenance personnel in mind; provide quick-disconnect, easy-to-assemble fittings and valves; avoid confined spaces for suspected problem areas; and enhance flow patterns or directions by proper use of tees, wyes, and elbows. Check valves should not be used, because slurry systems tend to cause them to plug and fail.

For example, a handy flushout within the piping network can be provided by placing a wye on its back, with the branch having a removal plug. This simple configuration provides easy flushing and cleaning without disassembling the piping.

Also, pipe reducers should be avoided, because hydraulic conditions induced by this fitting can cause "dewatering" (loss of fluidness) and compaction of the lime slurry.

Much has been written about the type of piping material to use. Slurry feed systems in the past have employed heavy-duty plastic pipe (Schedule 80), flexible rubber hoses, and either stainless or galvanized steel pipe. The final choice of a particular pipe material depends largely on personal preference, economics, anticipated problems, type of slaking water, and ease of assembly.

Perhaps the best piping network for a slurry system was seen in Pennsylvania at the Slippery Rock Acid Mine Drainage Plant constructed in 1967 under "Project Scarlift." It utilizes clear braided or reinforced tygon tubing connected to stationary fittings (pump inlets and outlets) with radiator clamps. This system allows constant flow observation, which immensely enhances location of blockages, along with easy breakdown for cleaning. Other advantages to this system are the ability to incorporate continuous long radius bends (ideal for slurry flows), and the ability of the pipe to flex under blows from a rubber mallet to loosen buildups.

The primary disadvantages are limitations in size, difficult procurement, and relative expense. The material cost can be overwhelmingly justified, however, by the savings in labor from maintenance.

A flushing system is not imperative in every situation. It is, however, highly recommended for any system undergoing frequent shutdowns (i.e., for any AMD plant that does not operate daily or operates only in wet weather).

Ideally, a minimum flow velocity of 1.0-1.2 m/s (3-4 ft/s) should be maintained within the slurry loop. Consequently, pipe sizing must be designed and not arbitrarily selected; piping should not be oversized because of anticipated future expansion. This common mistake can cause initial operational and maintenance problems amounting to a surprising cost.

The types of valves used in the slurry feed system determine, to a large extent, the degree of operation or maintenance. Among the many types available, pinch valves have proven most successful. The nature of their operation provides a self-cleaning mechanism. Pinch valves will close and perform efficiently, despite solid accumulations in the tube, and easily release any dewatered lime deposits upon opening.

Pinch valves are not foolproof, however, and can cause operational headaches if not properly sized (1). The tendency to use pinch valves the same size as the slurry piping has resulted in oversizing and poor valve operation. Port sleeves within a pinch valve are available in various sizes adaptable to operational flow conditions and should be designed properly.

Ball or plug valves can be used in the slurry piping system for either fully opened or fully closed conditions. The use of these valves for throttling implies obvious disadvantages and results in problems. These valves are best employed near pumps or on bottom nozzles of tanks.

A tight control on the metering devices (lime feeder, water valve) is imperative for the bleed-feed slurry system to work. Extreme variations in slurry concentrations (specific gravities) produce an inferior system with oscillating response times. A process designed to meet the demands of the control system by varying the slurry concentration generally WILL NOT work.

3.2.9 Slurry Tanks

Storage tanks for lime slurries can come in a variety of shapes and sizes. Generally, materials of construction can be whatever suits structural requirements and resists the high pH of the slurry. The configuration is usually dictated by the space available or assigned. Storage tank layout should tend to eliminate short-circuits in flow. This will help prevent unstabilized slurry from entering the treatment system. To reiterate, primary attention should be given to simplicity of transport between slurry preparation and storage tanks.

Tanks exceeding 9,500 l (2,500 gal) should be fitted with baffles, set 90° apart, to prevent vortex formation. Baffle widths equal to one-twelfth the tank diameter are sufficient in most cases (see Chapter 4, Mixing).

Slurry storage tanks with adequate agitation must be provided. Slurries resulting from proper lime slaking require relatively low-energy agitation to maintain the suspension. Hydrated lime particle size is small and its settling rate is slow. Mixer horsepower and impeller size and shape should be designed to keep impurities as well as lime in suspension. Otherwise, grit particles and precipitated salts might become troublesome sediment.

Agitator requirements are influenced by the tank size and shape. When available space allows some latitude in tank configuration, consult the agitator supplier for recommendations. These can direct the designer toward the most efficient application of his equipment.

3.2.10 Slurry Feed Control

At most plants, the treatment of AMD need not involve a sophisticated slurry feed control system. Usually the drainage flow is constant both in quantity and quality because of the beneficial effect of an equalization basin. The system is best controlled by the pH of the flow leaving the neutralization tank. This pH signal can control the dry lime feeder, a lime solution feeder, or a control valve. All three systems work well; but they are discussed here in order of preference.

Where drainage flow is fairly constant and the quality does not vary significantly, the dry lime feeder system can be adjusted so it operates constantly, and the lime slurry continuously overflows from the stabilization tank into the neutralization tank. This control method works well if ferrous iron concentrations are less than 100 mg/l. A 2.0- or 3.0-pH unit control range (usually 7.0-9.0) can be set on the controller, and any variance outside these limits should sound an alarm or stop the system until adjustments can be made. Periodic adjustments to the feeder or makeup water flow may be necessary to control pH trends. More than likely, this will be the result of gradual changes in the AMD quality or variations in the lime quality.

Where slakers discharge continuously to the stabilization tank, the chances are better that control can be maintained by the slurry feed rate. Response time is less, but slakers do not change their output at once when input is changed. Paste slakers perform better than slurry slakers in these circumstances because they maintain a constant paste concentration. Changes in output occur as the paste flow rate and the small dilution compartment concentration are changed. A slurry shaker must change the concentration of the entire slaker before a rate change is completed. The time for completion of such a concentration change may be too long and the system will shut down.

The importance of the slurry preparation method cannot be overstressed. Where short response time is required, meeting the demands of the control system by varying the slurry concentration WILL NOT work. The preparation of a constant slurry concentration and variation of the rate of application will work. The possible exception is a paste slaker or a hydrated lime feeder with varying feed control and small dilution tank from which the lime is sluiced directly to the neutralization tank.

The second type of feed system is a versatile type of solution feeder that has been used successfully for most types of lime suspensions. It is the dipper wheel type of feeder, sometimes called the Archimedes Wheel. This feeder, usually comprised of eight dippers that hold 0.5 l each, is rotated by a variable-speed drive. When submerged in a tank with a constant level, the dippers fill and discharge into an outlet trough. A mechanical counter driven by the shaft indicates dipper revolutions. It can be completely automated pneumatically, electrically, and mechanically with a ratio controller, or with the variable-speed drive and remote pH control (see Figure 3-10).

Another system, designed primarily for feeding lime slurries, is regulated by a pH probe and controller. It is supplied as part of a compact, complete package neutralization treatment system for permanent or portable installations. The whole unit measures 2.6 m x 1.6 m x 1.4 m (8.5 ft x 5.25 ft x 4.5 ft). It comprises a slurry mixing tank of either 0.6- or 0.9-m^3 (160- or 240-gal) capacity with a flash mixer; a pump, if gravity flow of the slurry is not possible; and a highly automated pH sensor at the discharge end of the unit, which regulates the flow of slurry into the treatment process. This system is based on using hydrated lime, either introduced manually by feeding 22.7-kg (50-lb) bags of lime into the mixing tank or metered from an overhead dry feeder. The lime is mixed into a uniform concentrated slurry.

The slurry flows by gravity or is pumped to a rotating cup wheel, where it is picked up and discharged to a funnel leading to the treatment tank. This funnel, located inside the wheel, is provided with an adjustable hood that controls the lime slurry feed. Control settings are provided by a controller and pH probe.

The feed rate is claimed to be infinitely variable over a 1-100 range, and the largest capacity model can feed up to 227 kg/hr (500 lb/hr) of hydrated lime in slurry form. To provide for continuous pH (and feed) control, a spare pH electrode is maintained in operating condition and used interchangeably.

The third type of lime slurry feeding involves proportioning the slurry flow in response to the pH signal. This most workable system involves pumping lime slurry through a continuous pipe loop from the stabilization tank and returning it to that tank. This pipeline should be sized so there is a significant head on the slurry circulating pump. The slurry is fed from a branch line and controlled by a pinch valve or another type of throttling valve. As previously mentioned, pinch valves are usually controlled by pneumatic systems. This system is best applied when there will be frequent variations in flow or quality, and when strict pH control is required.

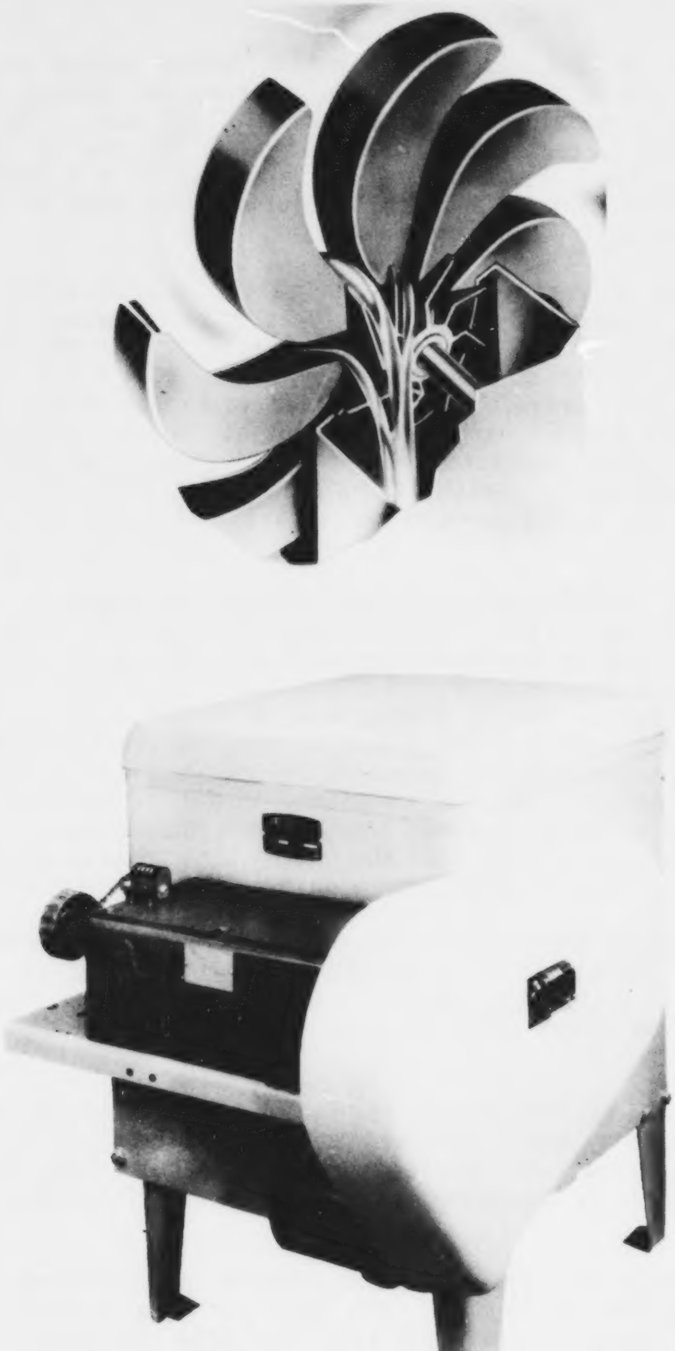


Figure 3-10. Dipper wheel and slurry feeder.

3.2.11 Pumps

Pumps for lime slurries generally fall into two categories: controlled volume and centrifugal. Controlled volume pumps are usually reciprocating diaphragm or progressive cavity types. The diaphragm metering pump has limited capacity and is not normally used in mine drainage treatment. Progressive cavity pumps are available in much higher capacities, but they wear faster than a centrifugal pump under similar operating conditions. An important consideration is maintaining adequate pipeline velocities with the modulating flow produced by these pumps. If the pumps do not provide adequate pipeline velocities at minimum rates, a recirculated slurry loop should be provided.

Centrifugal pumps and control valves are the obvious choices for a wide range of slurry flows. They are inexpensive, and standard designs incorporate flow patterns that lend themselves to easy slurry transfer. Two common configurations have been used with transfer pumps and different control valve operations (see Figures 3-11 and 3-12). Both methods work satisfactorily; therefore, they become a designer preference.

Manufacturers of centrifugal slurry pumps regard lime suspensions as fairly mild among slurries. These pumps are usually cast iron with replaceable liners and semiopen impellers. The designer should consider the tendency of lime to dewater under certain conditions of velocity and turbulence. Pumps should be easy to disassemble for maintenance and cleaning. Speed of rotation should not exceed 1,750 r/min (revolutions per minute), and should be as low as hydraulic requirements allow.

Although there appears to be a lack of agreement as to the kilowatts (horsepower) needed to pump lime slurries, the following formula is generally used:

$$\mathrm{kW} = \frac{Q \times W \times H}{76 \times E} \qquad \mathrm{hp} = \frac{Q \times W \times H}{550 \times E} \tag{5}$$

where kW = power in kilowatts	hp = power in horsepower
Q = liquid flow in m^3/s	Q = ft^3/s
W = specific weight of lime slurry in kg/m^3	W = lb/ft^3
H = head in m	H = ft
E = hydraulic efficiency of pump in decimal fraction	E = hydraulic efficiency of pump in decimal fraction
76 refers to kg-m	550 refers to ft-lb

For large pumps, the efficiency is in the 70%-80% range; but for small pumps, 40%-50% is generally used.

Pump shaft seals should be specially designed for slurries and should not be

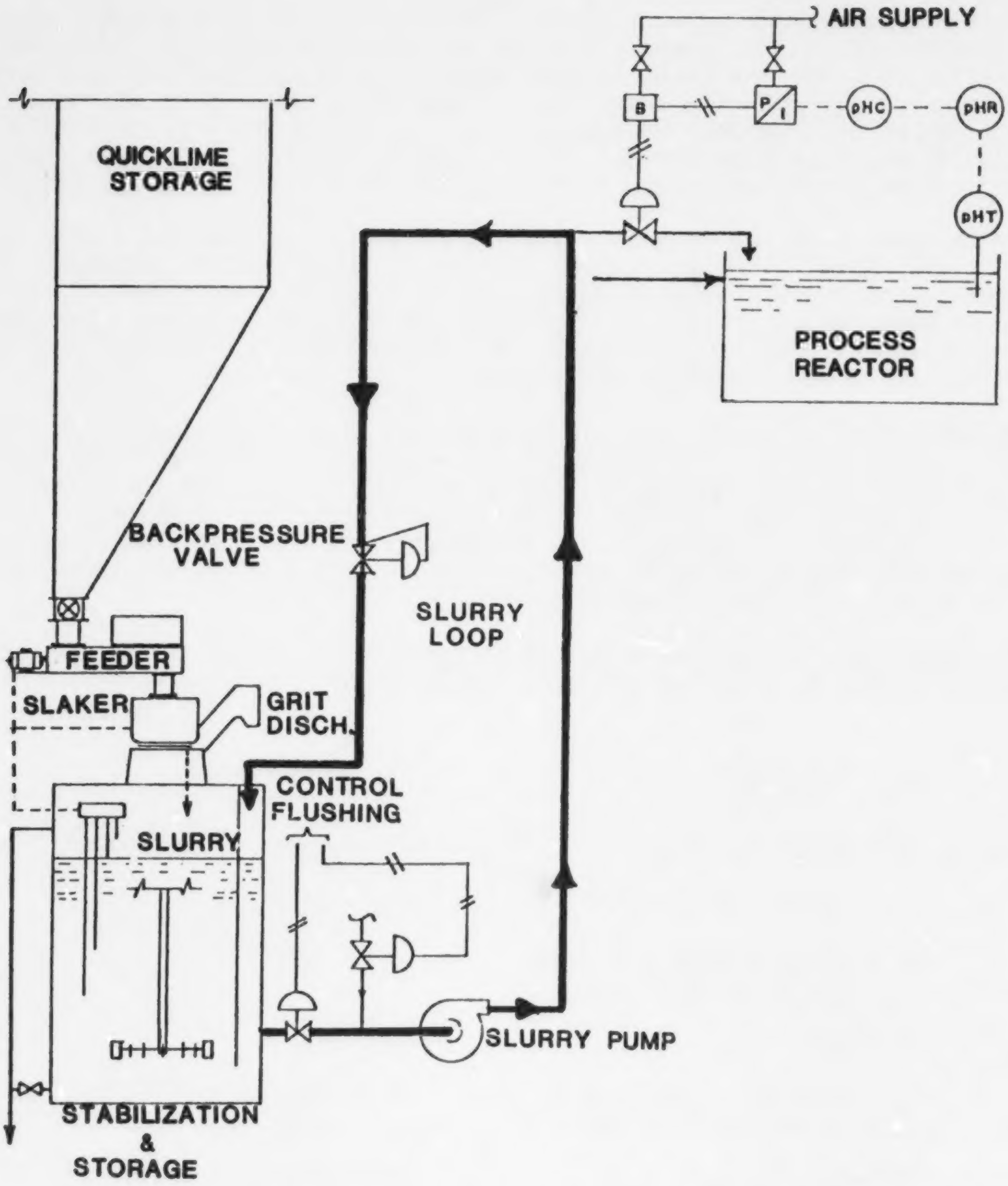

AIR SUPPLY
B
P/I
pHC
pHR
pHT
QUICKLIME STORAGE
PROCESS REACTOR
BACKPRESSURE VALVE
SLURRY LOOP
FEEDER
SLAKER
GRIT DISCH.
CONTROL FLUSHING
SLURRY
SLURRY PUMP
STABILIZATION & STORAGE

Figure 3-11. Slurry feed with pH control loop.

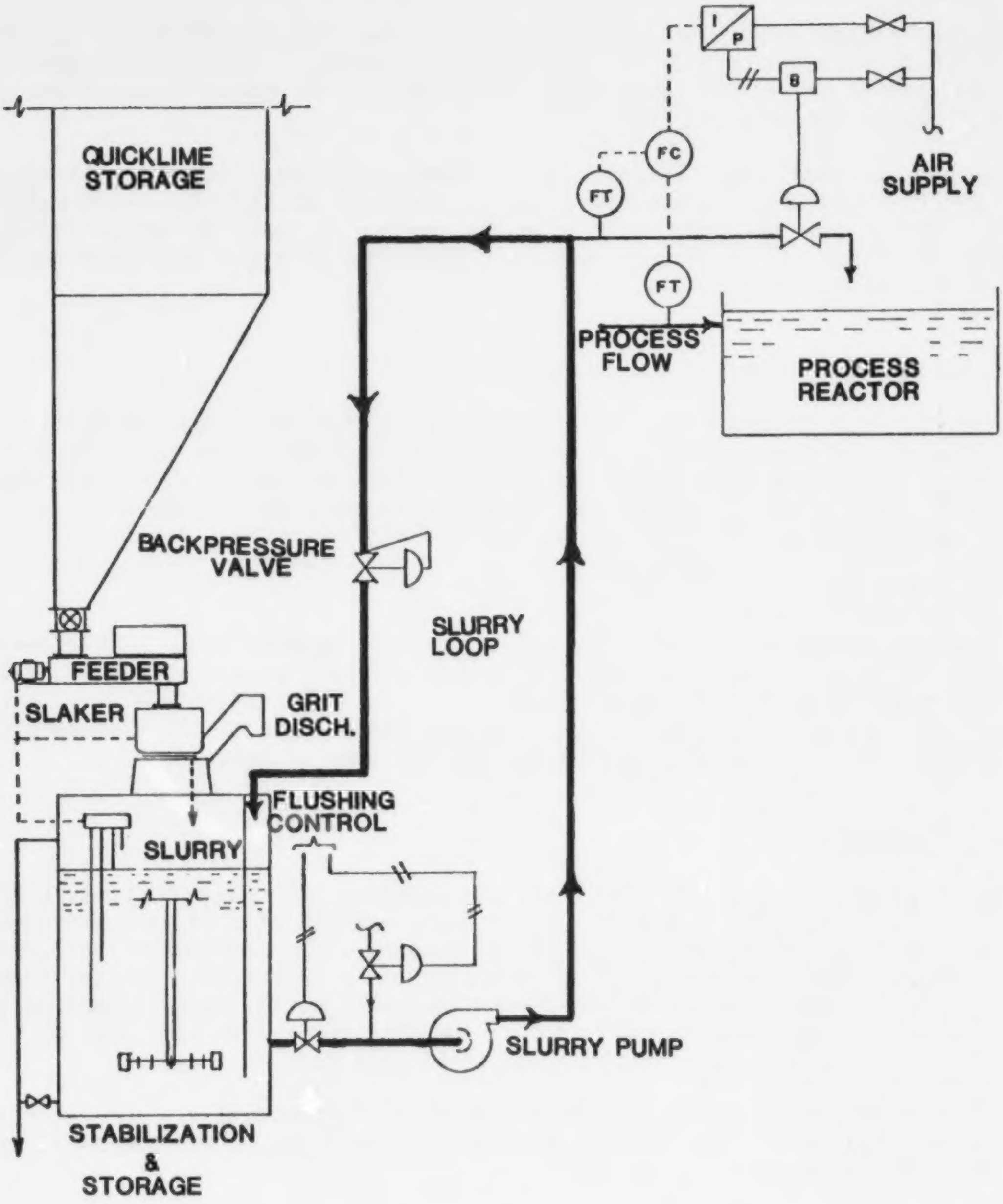

QUICKLIME STORAGE
I/P
B
AIR SUPPLY
FC
FT
FT
PROCESS FLOW
PROCESS REACTOR
BACKPRESSURE VALVE
SLURRY LOOP
FEEDER
SLAKER
GRIT DISCH.
FLUSHING CONTROL
SLURRY
SLURRY PUMP
STABILIZATION & STORAGE

Figure 3-12. Slurry feed by flow proportioning.

water-flushed. Introduction of water into the slurry will reactivate scaling. This usually will not occur in the pump, but starts to form shortly downstream. In any event, it can be avoided by using dry seals.

Pumps and piping should be sized for minimum required velocities in the recirculated slurry loop plus the process requirements. The specific gravity of the slurry should be taken into consideration in these calculations as well as those for valve sizing.

Standby pumps should be included in accordance with requirements for process reliability. If the process operation is critical to facility performance, full standby can be justified. Standby pumps, as well as other piping, should be placed in a manner that will minimize or avoid the settling of solids in the static sections.

3.2.12 Miscellaneous Considerations

Two of the most frequent questions presented to designers about lime slurry handling relate to friction loss and viscosity in pumps. For straight pipelines up to 152.4 m (500 ft) in length, the friction loss is relatively negligible. With several elbows and turns, there will be a greater friction loss, particularly with lime slurry containing some grit. In designing a small pipeline, it is common practice to increase the pump horsepower about 10% to allow for friction loss.

Viscosity of lime slurry is highly variable and depends upon such diverse factors as type of lime used (hydrate vs. quicklime), particle size (surface area), reactivity, slaking water, lime ratio, initial temperature of slaking water, slurry concentration, and temperature. Nevertheless, the viscosity of a usable slurry (10% solids) can be assumed the same as water.

3.3 Limestone

There has been considerable research, development, and demonstration in the use of limestone for the neutralization and treatment of acid mine drainage. On the surface, there appears to be an economic advantage with limestone, which is available at about 30% of the cost of quicklime or hydrated lime. As will be explained, there are inherent disadvantages for using limestone in this kind of treatment process. As a result, there are very few if any operating facilities using limestone to treat acid mine drainage.

The dolomitic (magnesium) and high-calcium form of limestone are available in a variety of sizes. Commercially available sizes range from 76.2 mm (3 in) to 0.074 mm (200 mesh).

Although there are two types of limestone, the high-calcium form has undergone the most study as a neutralizing agent because of better reactivity. The rate of reaction for dolomitic limestone was studied equally but found to be ineffective. Subsequent discussions about limestone refer only to the high-calcium form.

High-calcium limestone usually has the following chemical composition (1, 2):

	Calcium Limestone (Rock Dust) % Composition
Calcium Oxide (CaO)	53.0 - 56.0
Magnesium Oxide (MgO)	0.12 - 3.11
Calcium Carbonate ($CaCO_3$)	92.66 - 98.6
Silica Dioxide (SiO_2)	0.1 - 2.89

The reaction of limestone with an acid is accompanied by the liberation of carbon dioxide (CO_2), expressed by the following equations.

$$\text{Limestone} + \text{Strong Acid} \rightarrow \text{Gypsum} + \text{Carbonic Acid}$$

$$CaCO_3 + H_2SO_4 \rightarrow CaSO_4 + H_2CO_3 \qquad (6)$$

$$3CaCO_3 + Fe_2(SO_4)_3 + 6H_2O \rightarrow 2CaSO_4 + 2Fe(OH)_3 + 3H_2CO_3 \qquad (7)$$

$$3CaCO_3 + Al_2(SO_4)_3 + 6H_2O \rightarrow 3CaSO_4 + 2Al(OH)_3 + 3H_2CO_3 \qquad (8)$$

The effervescent reaction, or release of carbon dioxide, from limestone neutralization is not so obvious with acid mine drainage as with stronger acids, but it does occur.

To utilize limestone effectively as a neutralizing agent, certain quality criteria must be maintained. An early study by Bituminous Coal Research (BCR) involving two years of investigation concluded that the effectiveness of limestone as a neutralizing agent depends upon the following criteria (3):

1. minimum particle size, preferably a minus 325 mesh (approximately 0.044 mm);
2. high calcium content, approaching pure calcium carbonate;
3. low magnesium content, thus eliminating dolomites ($CaMg(CO_3)_2$) and magnesites ($MgCO_3$);
4. high specific (surface) area.

There are many different quality limestones available; however, finding a limestone that meets all these criteria can be difficult.

3.3.1 Treatment of Ferrous Iron Drainage

An EPA study by Wilmoth (1) indicates that limestone treatment of ferrous iron streams is possible, but economically undesirable. The most significant factors inhibiting full-scale use of powdered limestone for AMD treatment are its slow reactivity and its inability to increase the pH above 7.0 (maximum obtainable was 7.4). A pH above 7.0 could be achieved only through effluent recycle and increased reaction time, which enlarges the process units.

At pH 7.0, the slow rate of oxidation of ferrous iron is a deterrent. Excessive aeration times of several hours (normally 30 minutes or less at pH 8.5) are needed to oxidize the ferrous iron completely. In addition to increasing the normal costs for the aeration unit, the longer detention periods have a tendency to destroy the floc particles required for good settling, which results in an effluent with high suspended solids and turbidity.

According to Wilmoth, the final process scheme that produced a satisfactory effluent incorporated a 20% sludge recycle rate to maintain a pH near 7.4; 30 minutes detention within the flash mixer; 4-6 hours aeration; and finally, coagulant addition for reflocculation and effective settling of the sludge and limestone particles.

The reagent cost alone for this process approached \$0.15/3.8 m^3 (1,000 gal), which essentially nullifies any economic advantage in using limestone over other lime forms. The additional capital investment for enlarged process units (i.e., flash mixer, aerators, and aeration tank) and power costs totally invalidate the practicality of the limestone treatment system for high ferrous iron drainages.

Perhaps a better design criteria (3) for limestone treatment of AMD would be:

Ferrous Iron Concentration (mg/l)	Response to Limestone Treatment
0-50	Effective treatment may be achieved without pre- or postneutralization iron oxidation.
50-100	May be effectively treated but requires postneutralization aeration and significant reaction-retention time. Preneutralization oxidation can reduce ferrous iron concentrations.
> 100	Potential treatment is uncertain with experience to date, unless combined with preneutralization ferrous iron oxidation to achieve the above ferrous levels.

Although only a guide, these statements are "cautionary" and imply skepticism. More recent studies supplement these statements further by discouraging the use of limestone with water having more than 100 mg/l ferrous iron.

Lovell also emphasizes many significant design parameters that must be evaluated before selecting a limestone neutralization process (4). Those highlighted are:

1. specifications for limestone grade, size, and hardness;

2. mode of operation (mixer or rotary mill);
3. gas exchange capabilities (oxidation and CO_2 removal);
4. supplementary reagent requirements (lime, polymers);
5. operating pH and aeration requirements for ferrous iron oxidation;
6. ratio of recycle volume;
7. sludge settleability.

3.3.2 Treatment of Ferric Iron Drainage

To this point, only the feasibility of successful limestone treatment of mine drainage containing ferrous iron has been discussed. If iron is mostly in the ferric form (at least a 4:1 ferric-to-ferrous ratio), treatment with limestone appears feasible. Wilmoth and Hill, however, could only realize a 32% limestone utilization efficiency in their studies on Grassy Run drainage at Norton, W. Va. (5). Consequently, it would take three times the amount of limestone to neutralize this water. At the time of the study (1970), limestone cost less than one-third that of lime, which placed it at an economic disadvantage. Today, with the price (including delivery) for limestone at \$24/Mg (\$22/ton) and hydrated lime at \$72/Mg (\$65/ton), the treatment economics are more favorable.

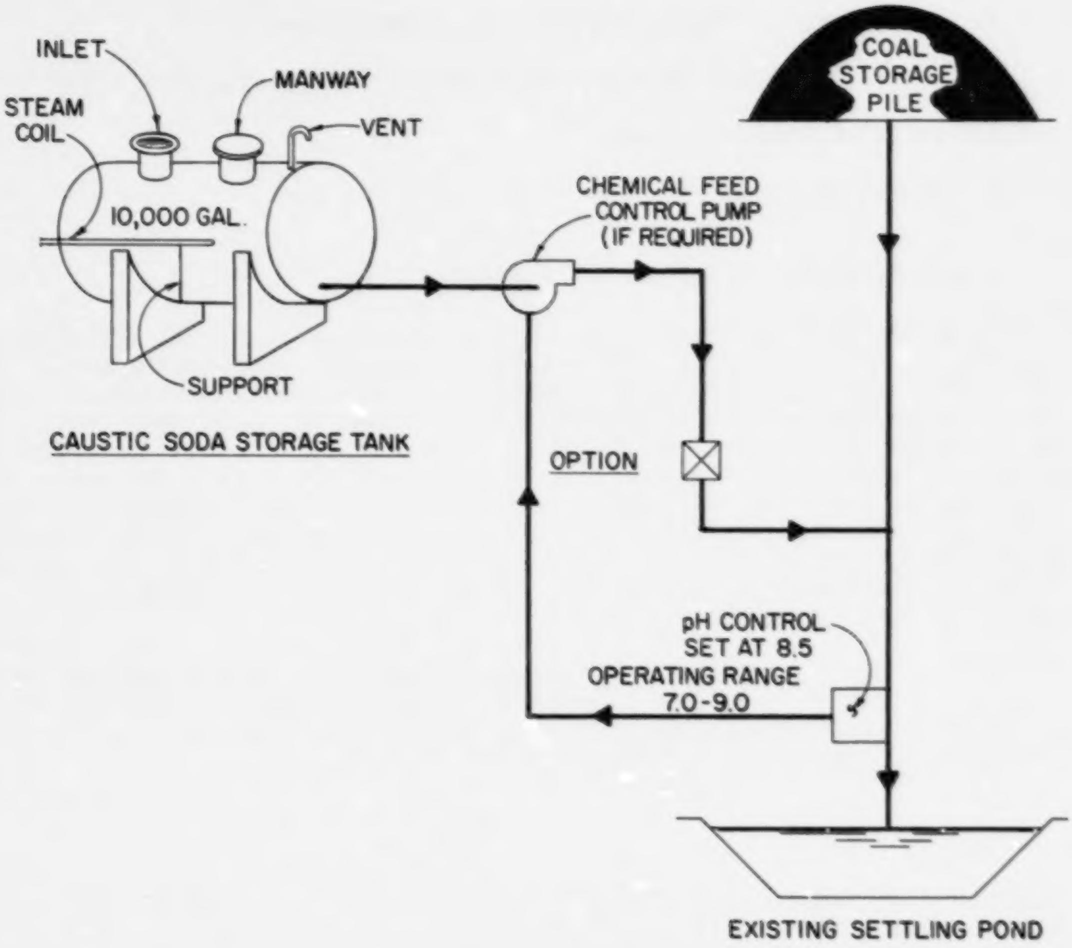
In summary, the use of limestone for the treatment of AMD has decisive disadvantages, both economic and functional. Limestone cannot be used to treat drainages containing manganese because the maximum obtainable pH is not sufficient to precipitate the manganese.

3.4 Caustic Soda

Traditionally, caustic soda (NaOH) has been employed as a neutralizing agent for low-flow, mildly acidic drainages located in remote areas. Generally associated with surface mine drainage problems, it is extensively used for neutralization in the panhandle region near West Virginia, Ohio, and Pennsylvania.

The conventional process usually consists of a horizontally mounted 38-m^3 (10,000-gal) storage tank for the caustic soda, and a flume-type chemical feeder (Figure 3-13). Gravity serves as the source of power, with a constant head valve in the chemical feeder controlling a constant feed.

In surface mining operations, caustic is frequently used to neutralize acid water as it is being pumped from the pit. For such a temporary or portable need, caustic can be siphoned into the suction side of a portable mine water pump as shown in Figure 3-14. The flow of caustic is controlled by adjustment of the gate valve on the storage tank outlet. The discharge line of the pump should be of sufficient length to provide 1-2 minutes retention for ade-


INLET
MANWAY
STEAM COIL
VENT
10,000 GAL.
SUPPORT
CAUSTIC SODA STORAGE TANK
CHEMICAL FEED CONTROL PUMP (IF REQUIRED)
COAL STORAGE PILE
OPTION
pH CONTROL SET AT 8.5
OPERATING RANGE 7.0-9.0
EXISTING SETTLING POND

50% Caustic Feed Options

1. Variable chemical feed pump with pH probe controller.
2. Solenoid valve with pH probe controller (gravity).
3. Constant chemical feed pump with electric pinch valve and pH probe control.

Figure 3-13. Caustic soda treatment system.

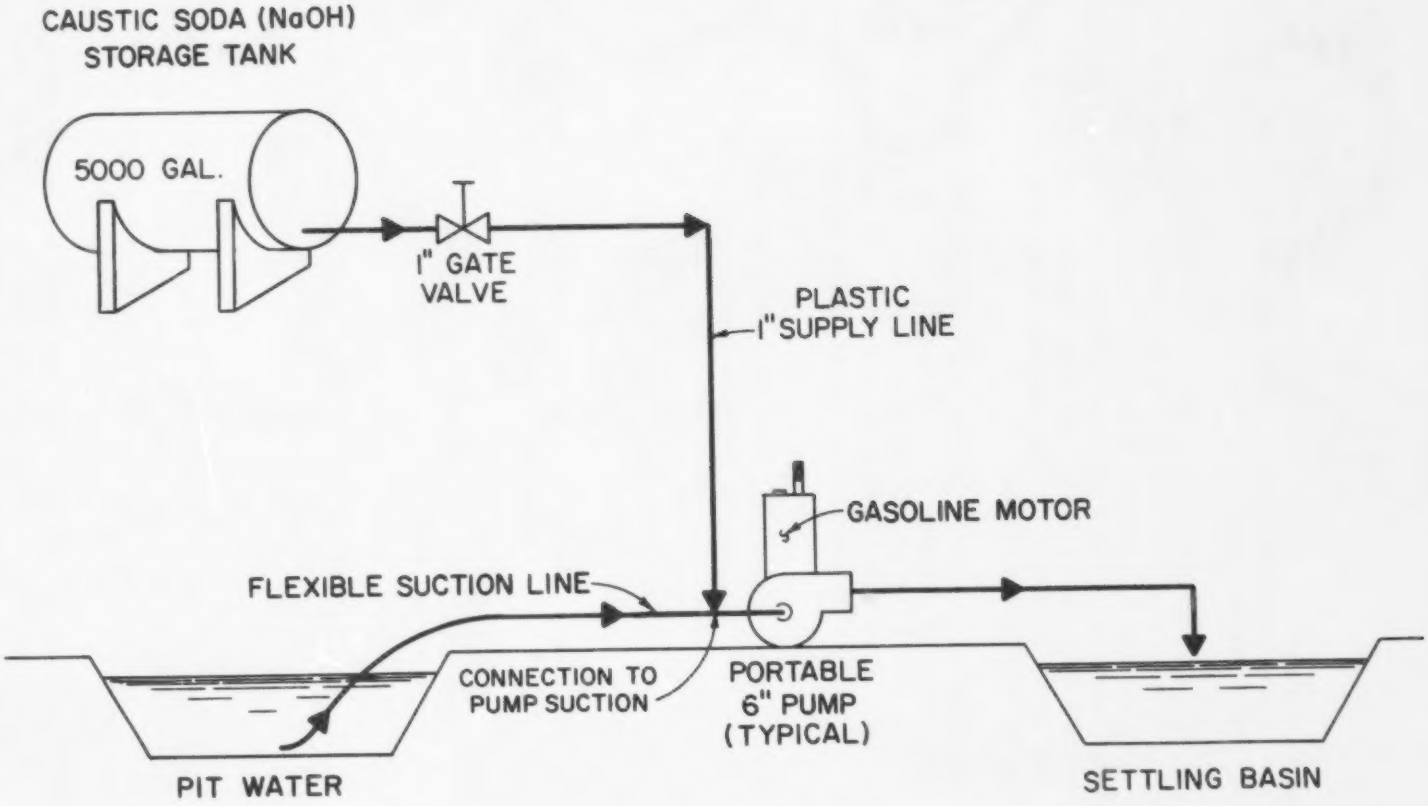

CAUSTIC SODA (NaOH)
STORAGE TANK
5000 GAL.
1" GATE VALVE
PLASTIC 1" SUPPLY LINE
GASOLINE MOTOR
FLEXIBLE SUCTION LINE
CONNECTION TO PUMP SUCTION
PORTABLE 6" PUMP (TYPICAL)
PIT WATER
SETTLING BASIN

Figure 3-14. Portable caustic soda feed arrangement.

quate mixing of the caustic with the mine drainage.

The decisive limiting factor to the use of caustic soda in AMD treatment is its cost. Caustic soda costs average about $287/dry kkg ($260/dry ton). Since it is usually bought as a liquid (50% concentration), the purchaser pays for hauling 50% water, which further increases this cost. In cold weather, caustic is purchased as a 20% solution, which allows for the lowest temperature before freezing occurs.

Aside from the dangers in handling caustic, the chemical characteristics of concentrated solutions require design consideration. Table 3-2 illustrates some of the properties of sodium hydroxide solutions. This knowledge is extremely helpful when one also knows the theoretical alkali requirements. A close approximation for the dilution of the concentrated solution (50%) can be computed and made directly at the storage area upon delivery, following all appropriate safety precautions.

Figure 3-15 presents the freezing points for various solution strengths. The 50% caustic soda solution freezes at 12°C (54°F), and is quite viscous (barely pumpable) at a slightly higher temperature of 15°C (60°F). Ideally, an 18% solution offers the lowest freezing temperature; however, higher concentrations have been used in coal-producing states with minimal freezing problems in the winter.

Despite its detrimental properties, sodium hydroxide will produce an excellent effluent quality. A properly treated drainage will be low in suspended solids and turbidity, and will have an iron content within limitations. Also, caustic soda is nearly 100% reactive. Generally, sodium hydroxide sludges are fluffy and more voluminous than lime sludges; however, they display acceptable settling properties (5).

In his experiments at the Crown Mine drainage field site, Kennedy encountered several problems related to winter weather that must be overcome for caustic soda to be reliable (6). One of these was acquiring a good water supply to dilute the sodium hydroxide. This can be a severe problem. An illustration of the type of chemical feeder used by Kennedy is shown in Figure 3-16. Devices such as these are available from several manufacturers with installed costs in the area of $10,000.

The settling ponds in many of these facilites are merely depressions in reclaimed strip land. The volume of sludge associated with this type of treatment will vary from 4% to 6% of the flow, depending upon the acidity and iron content of the drainage.

The use of sodium hydroxide for treatment of high sulfate drainages (greater than 2,500 mg/l) has merit because there is no gypsum ($CaSO_4$) formation. The effluent, however, will have a higher dissolved solids content because of the total solubility of sodium sulfate.

In summary, caustic soda is an excellent neutralizing agent capable of producing an effluent within discharge limitations. The high cost of this reagent, along with its undesirable handling properties, limits its use.

TABLE 3-2

DENSITY OF AQUEOUS SODIUM HYDROXIDE SOLUTIONS AT 20°/4° C VALUES GIVEN IN THE INTERNATIONAL CRITICAL TABLES

	Weight of NaOH in Solution					
Specific Gravity	Grams per liter	Pounds per U.S. gallon	Pounds per cubic foot	Percent NaOH	Degrees Baume	Degrees Twaddell
1.0095	10.10	0.08425	0.6302	1	1.4	1.90
1.0207	20.41	0.1704	1.274	2	2.9	4.14
1.0318	30.95	0.2583	1.932	3	4.5	6.36
1.0428	41.71	0.3481	2.604	4	6.0	8.56
1.0538	52.69	0.4397	3.289	5	7.4	10.76
1.0648	63.89	0.5332	3.989	6	8.8	12.96
1.0753	75.31	0.6285	4.701	7	10.2	15.16
1.0869	86.95	0.7256	5.428	8	11.6	17.38
1.0979	98.81	0.8246	6.169	9	12.9	19.53
1.1089	110.9	0.9254	6.923	10	14.2	21.78
1.1309	135.7	1.133	8.472	12	16.8	26.18
1.1530	161.4	1.347	10.08	14	19.2	30.60
1.1751	188.0	1.569	11.74	16	21.6	35.02
1.1972	215.5	1.798	13.45	18	23.9	39.44
1.2191	243.8	2.035	15.22	20	26.1	43.82
1.2411	273.0	2.279	17.05	22	28.2	48.22
1.2629	303.1	2.529	18.92	24	30.2	52.58
1.2848	334.0	2.788	20.85	26	32.1	56.96
1.3054	365.8	3.053	22.84	28	34.0	61.28
1.3279	398.4	3.325	24.87	30	35.8	65.58
1.3490	431.7	3.603	26.95	32	37.5	69.80
1.3696	465.7	3.886	29.07	34	39.1	73.92
1.3900	500.4	4.176	31.24	36	40.7	78.00
1.4101	535.8	4.472	33.45	38	42.2	82.02
1.4300	572.0	4.774	35.71	40	43.6	86.00
1.4494	608.7	5.080	38.00	42	45.0	89.88
1.4685	646.1	5.392	40.34	44	46.3	93.70
1.4873	684.2	5.710	42.71	46	47.5	97.46
1.5065	723.1	6.035	45.14	48	48.8	101.30
1.5253	762.7	6.365	47.61	50	49.9	105.06

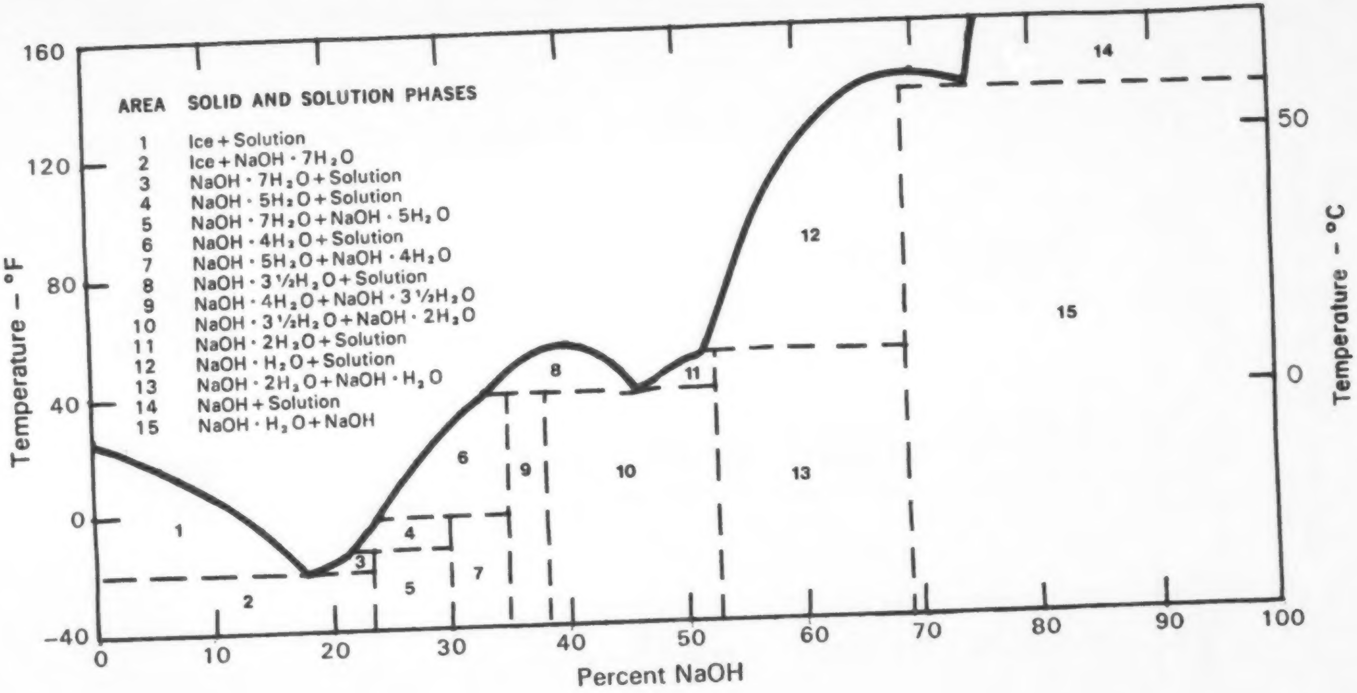


Figure 3-15. Freezing points of caustic soda solutions.

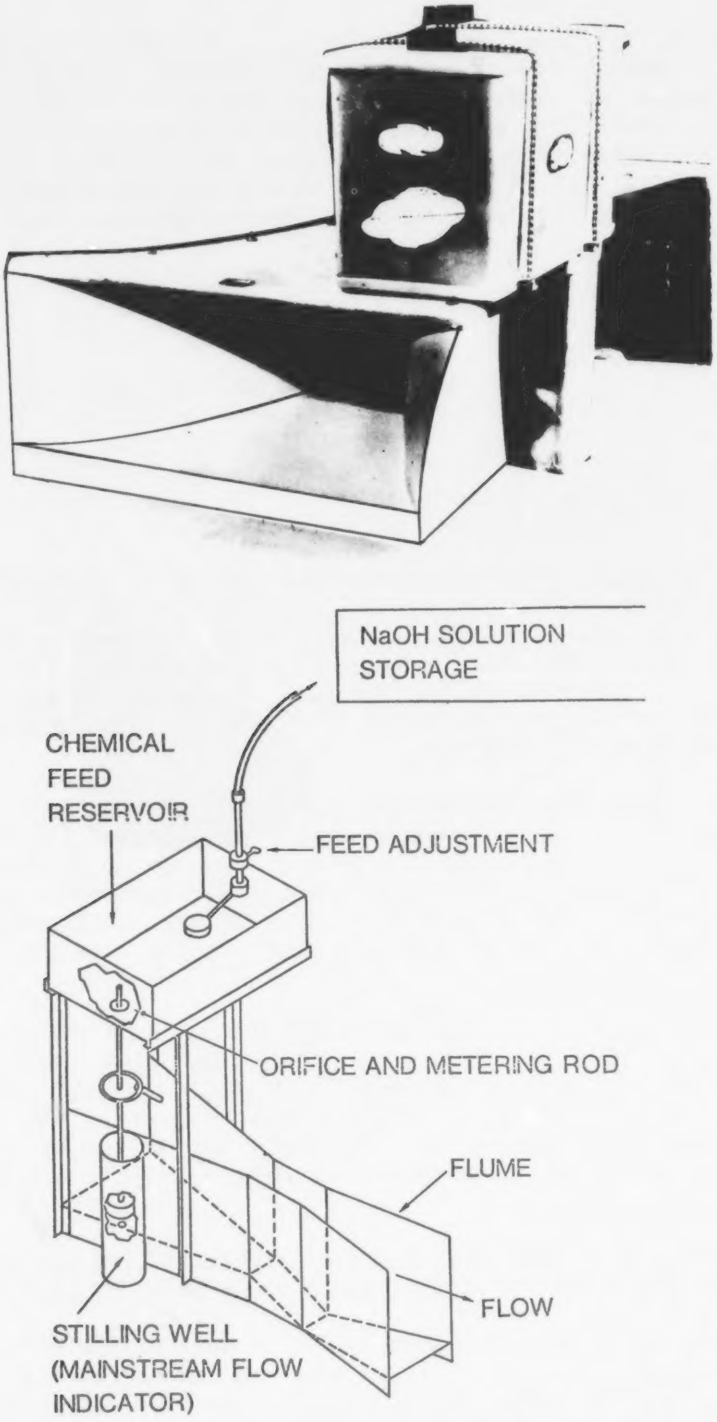

NaOH SOLUTION STORAGE
CHEMICAL FEED RESERVOIR
FEED ADJUSTMENT
ORIFICE AND METERING ROD
FLUME
FLOW
STILLING WELL (MAINSTREAM FLOW INDICATOR)

Figure 3-16. Flume chemical feeder.

Sodium hydroxide has been successfully used in remote locations to treat mildly acid or small flow drainages. The automatic feeding methods always present the danger of overtreatment in incidences of a feeder malfunction. Winter weather protection must be provided to minimize freezing problems. Finally, sludge produced by treatment with this chemical will have good settling properties, but it tends to be fluffy and more susceptible to washout. In most cases, the settling pond should represent the final disposal site for sludge.

3.5 Soda Ash

Soda ash (Na_2CO_3) has rarely, if ever, been used as the principal neutralizing agent in large-flow mine drainage treatment facilities. Due to its high cost of $309/Mg ($280/ton) and limited availability, soda ash is usually used only for treatment of low-flow drainages that contain little ferrous iron such as would occur in surface mines. Its selection for such applications is more for convenience than cost efficiency.

Soda ash is produced in solid pellet form called briquettes or "prills." Production of these is limited and the availability is scarce. Effective treatment has been demonstrated by simply immersing these prills in a wire basket in the flowing drainage. The rate of dissolution will enable such a system to be effective for a 24-hour period. The volume of drainage flow and its acidity will define the number of prills to be used each day. This is best determined by experimentation.

Devices are available for feeding soda ash prills over a longer period. One method consists of a storage hopper mounted over a basket (see Figure 3-17). The hopper usually accommodates one or two bags up to 90 kg (200 lb) of prills. The prills will tumble through the hopper opening at the bottom and keep the basket full. While this is a simple system, it is subject to clogging at the hopper opening as the soda ash absorbs moisture, causing the prills to expand.

This method of treatment is controlled by the rate of dissolution of the prills. Depending upon the acidity in the drainage, the final treated water may not be within the desired pH range. It should also be recognized that drainages with significant iron concentrations may cause the prills to be coated, rendering them ineffective. Multiple systems could be employed in series or parallel if one unit cannot adequately treat the flow or acidity involved.

In practice today, a typical soda ash small-scale feeding operation is shown in Figure 3-18. The cost for such a feeder would be about $5,000.

A more effective method of using soda ash is in conjunction with a dissolver, if electric power is available. Flaked soda ash can be fed into a batch tank (comparable to a slurry tank) equipped with a mixer to dissolve the soda ash in a water solution. The tank size would depend upon the amount of alkali (soda ash) needed, the time period involved, and the concentration of solution used. The soda ash solution would then be fed into the AMD stream by

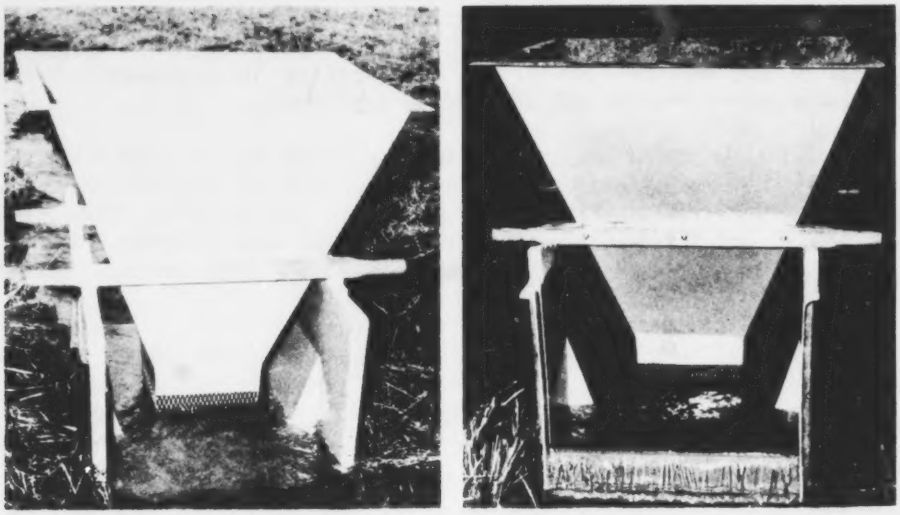


Figure 3-17. Soda ash prill hopper.

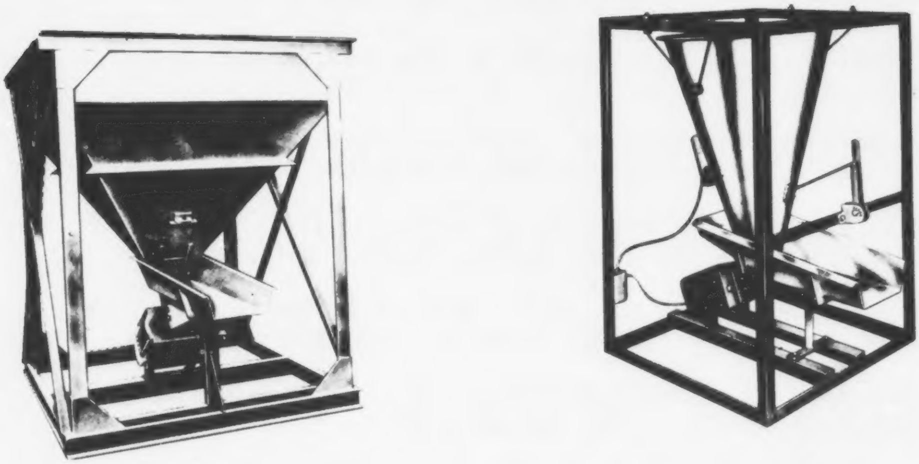


Figure 3-18. Soda ash vibrating feeders.

gravity with flow controlled by a valve, or pumped at a constant to variable rate depending on the degree of sophistication used (refer to Figure 3-13). This system offers better control and reliability in treatment; however, it does require a power source and additional equipment.

Wilmoth and Hill, in their studies on neutralizing ferric iron mine drainages (approximately 125 mg/l) with soda ash, were able to produce an effluent of satisfactory discharge quality, but treatment costs were impractical (5). Based on their results and today's prices, soda ash treatment would be the most expensive method for treating mine drainage and appears economically undesirable.

In laboratory tests with several mine drainages, Lovell found soda ash consumption ranged from 150% to 310% more than the theoretical requirement for neutralization (4). He explicitly reports a 56% use-efficiency in treating AMD, which means that twice the theoretical soda ash requirement will be required.

Both researchers found that the sludge formed by soda ash neutralization settled well and compacted to densities comparable to those obtained wth quicklime and hydrated lime.

3.6 References

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CHAPTER 4

MIXING

4.1 Introduction

Unit processes that essentially mix are incorporated into every acid mine drainage treatment facility. Mixing involves the uniform dispersion or suspension of liquid or solid particles into another liquid or solid media. Impeller design, rotational speed, and horsepower requirements vary according to the mixing application and the properties of the reagents.

Rapid mixing is usually employed in AMD treatment for the addition of the neutralizing agent to the raw water. Mixing is also required for preparation of lime slurries and polymer solutions. Gentle or slow mixing is associated with flocculation, large lime slurry storage tanks, and sludge thickening.

The purpose of this chapter is to make the designer aware of mixing technology and the techniques employed in sizing a mixer. The basic engineering principles involved with the proper sizing and design of the mixing vessel are also presented.

4.2 Types of Mixers

Mixers are classified into two broad categories according to the flow patterns they produce. Axial and radial patterns are general terms describing the kinds of mixing devices that impel or move the entraining media. Axial flow includes propellers, axial turbines, fan turbines, and pitched paddles. The axial and fan turbines are the most common. These mixing devices induce flow patterns parallel to the drive shaft. Radial flow impellers, on the other hand, induce currents perpendicular to the drive shaft and include radial turbines and paddles. Figure 4-1 illustrates the simple difference between the impellers that produce these two distinct flow patterns.

As shown in Figure 4-1, the basic difference between the two types of mixers is that the axial turbine blades are pitched. Consequently, each impeller type has its own characteristic flow patterns, making each more suitable for certain mixing applications. In this manual, emphasis will be on axial impellers because most mixing applications in AMD treatment require this type.

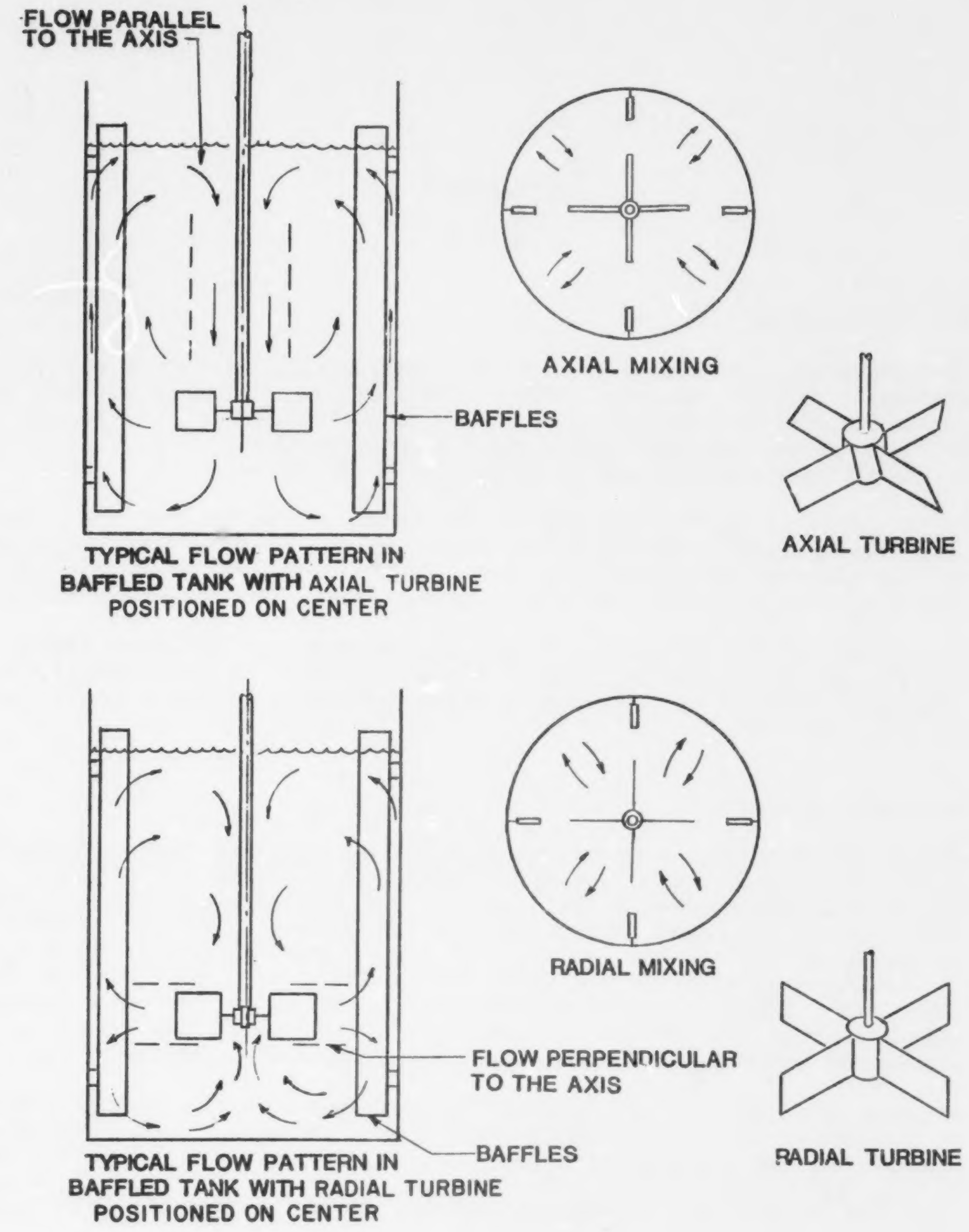

FLOW PARALLEL TO THE AXIS
AXIAL MIXING
BAFFLES
AXIAL TURBINE
TYPICAL FLOW PATTERN IN BAFFLED TANK WITH AXIAL TURBINE POSITIONED ON CENTER
RADIAL MIXING
FLOW PERPENDICULAR TO THE AXIS
BAFFLES
RADIAL TURBINE
TYPICAL FLOW PATTERN IN BAFFLED TANK WITH RADIAL TURBINE POSITIONED ON CENTER

Figure 4-1. Comparison of axial and radial flow patterns.

4.2.1 Propeller Mixers

Propeller mixers are used primarily for rapid mixing. Axial flow propeller mixers can be portable or fixed-mounted, depending on the mixer size and its application. Portable mixers are usually mounted on the side of the mixing vessel. Generally, angular, top-entering propeller mixers range in size from 0.37 to 2.24 kW (0.5 to 3.0 hp), although many designs limit the size to 0.75 kW (1.0 hp) and a maximum shaft length of 1.83 m (6.0 ft).

Portable mixers are usually mounted angularly off-center to obtain a top-to-bottom mixing pattern. This prevents a swirling pattern from developing and also eliminates the need for tank baffles. Typically, the maximum vessel volume used with angular mixers 2.24 kW (3.0 hp) or less is 3.785 m^3 (1,000 gal), but can vary with the application. The mixer shaft should enter at a 15° angle from vertical and, if possible, at a point off the tank centerline as shown in Figure 4-2.

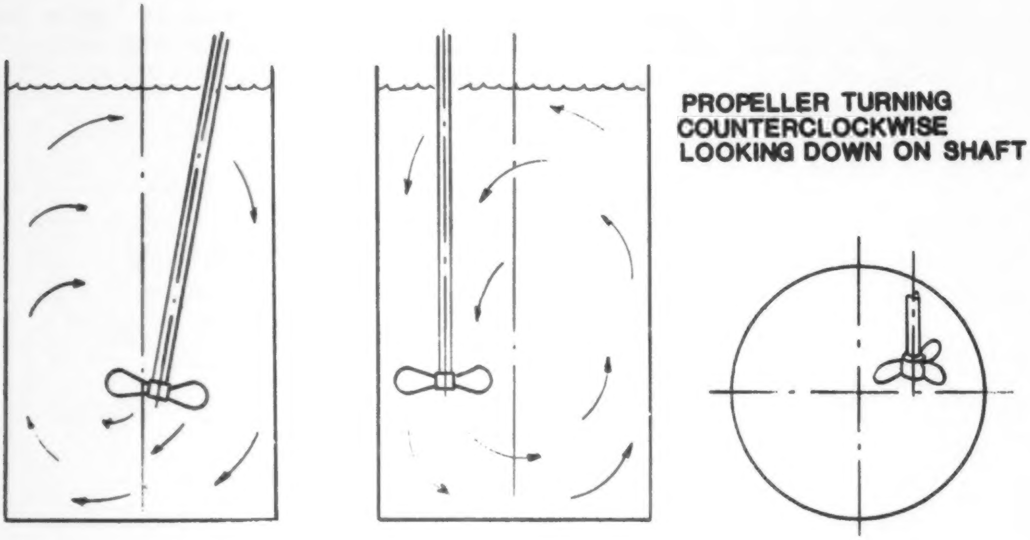


Figure 4-2. Off-center, top-entering propeller positions.

Fixed-mounted, right-angle drive propeller mixers should be positioned vertically, on-center, in a baffled tank.

Propellers are usually no larger than 46 cm (18 in) in diameter regardless of the vessel size. In deep tanks with on-center, top-mounted, right-angle, or vertical mixers, multiple propellers can be placed on a single shaft, usually aiming the liquid in the same direction.

The flow patterns produced by propeller and other axial mixers drive the water down to the tank bottom, then horizontally until turning upwards at the wall, and eventually return it to the suction side of the propeller (oval loop), as illustrated in Figure 4-1.

There are two basic speed ranges applicable to both portable and fixed-mounted propeller mixers. Direct-drive propellers (high speed) rotate at either 1,150 or 1,750 r/min, while gear-driven propellers (low speed) rotate within the 350-420 r/min range (1). The faster speeds provide a high level of shear with a low draft capacity, making it suitable for flash mixing of chemicals. The low speeds provide less fluid shear with a large draft capacity, making them more suitable for solids suspension applications, such as slurry makeup tanks less than 11.355 m^3 (3,000 gal) in size (2).

4.2.2 Turbine Mixers

Turbine mixers can be axial or radial flow, depending on the impeller design. Axial impellers are pitched-blade or fan turbines, while radial impellers are flat, curved, or with a spiral backswept blade as illustrated in Figure 4-3. The curved and spiral backswept impellers are used only for high-viscosity applications, which are not often encountered in AMD treatment unless mixing large quantities of sodium hydroxide or soda ash.

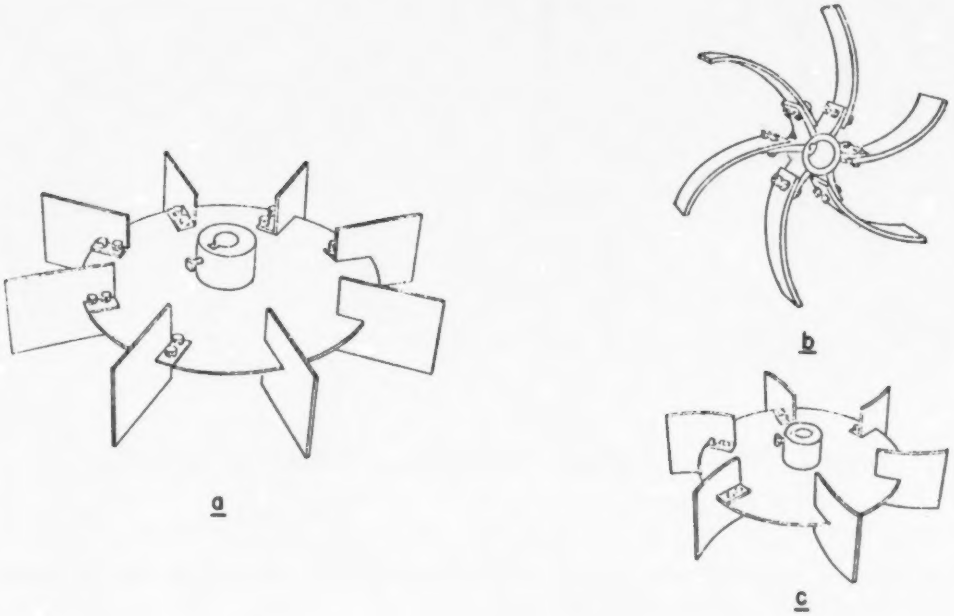
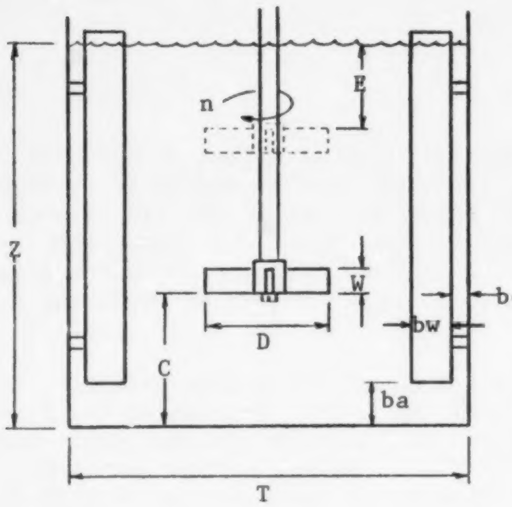


Figure 4-3. Typical radial turbine impellers: (1) flat-blade turbine, (2) spiral backswept turbine, (3) curved-blade turbine.

Axial turbines are used for most large-scale mixing applications involving liquid-solid suspensions such as mixing large, stored volumes of a lime slurry. Turbine mixers are usually fixed-mounted, vertically, in fully baffled tanks. This configuration gives good top-to-bottom fluid circulation throughout the mixing vessel. Turbine impeller diameters are usually one-third of the tank diameter, but can range between 30% and 40%.

The standard number of blades on a turbine impeller is either 6 or 8, but there can be anywhere from 4 to 16. The turbine impeller is mounted about 1 turbine diameter above the tank bottom. Turbine units are always mounted vertically. Their power range can be provided between 0.75 and 373 kW (1 and 500 hp). Common turbine impeller speeds are available between 50 and 150 r/min, but can be obtained in a range from 15 to 420 r/min (1).

When mounted in a tank with a liquid depth equal to or greater than its diameter, one turbine impeller is sufficient. If the liquid depth-to-tank diameter ratio exceeds 1:3, two turbine impellers must be used. The upper turbine should be located 0.5-1.0 turbine diameter below the liquid surface (E). The lower turbine should be mounted 1 turbine diameter above the tank bottom (C) as shown in Figure 4-4.



ba = Baffle Bottom Clearance
bc = Baffle Clearance
bw = Baffle Width
C = Turbine Bottom Clearance
D = Turbine Diameter
E = Upper Turbine Depth
n = Rotation Speed
T = Tank Diameter
W = Turbine Blade Height
Z = Liquid Depth

Figure 4-4. Characteristics of a mixing tank and standard turbine.

4.3 Baffles

Baffles are used with turbine impellers and on-center vertically mounted propeller mixers. The standard width for baffles (bw) used with turbine mixers is one-twelfth the tank diameter, while the baffle width used with propeller mixers is one-eighteenth the tank diameter (1). The standard number of baffles is four, mounted 90° apart. To prevent the formation of dead spots, the baffles should be mounted with a clearance between the baffles and the side wall (bc). This space should be 10%-15% of the baffle width; however, this clearance ranges between 2.54 and 7.6 cm (1 and 3 in). For tanks with flat or slightly conical bottoms, the baffles should end a minimum distance of one-half baffle width above the tank bottom (ba) (3). Also, baffles should extend at least 15.2 cm (6 in) above the liquid level.

4.4 Shafts and Drives

The small, high-speed turbine impellers, used with portable mixers, are connected directly to the drive motor. Slow-speed turbines have a shaft coupled directly to a speed reducer, in which case torsional bending stresses are transmitted to the reduction gears. The shaft can be isolated from the speed reducer by independent bearings. This provides a flexible coupling that does not transmit stresses to the speed reduction gears.

Shaft lengths are a function of the tank depth. Short shafts and those up to 4.57 m (15 ft) in length usually need no support, but longer shafts usually require foot bearings. These should be avoided if at all possible. Closed tanks require mechanical seals or stuffing boxes.

4.5 Energy Requirements

A general rule often used by designers for sizing mixer motors is 0.2 kW/m^3(1 hp/1,000 gal) of tank volume (2). Using the ideal turbine design as a starting point, the mixer speed (n) and exact turbine diameter (D) can be determined for the given slurry concentration in the tank. This procedure involves applying correction factors to the turbine diameter for differences between the ideal turbine and the one being designed in turbine depth and the specific gravity of the slurry (3).

The power dissipated during mixing is a function of the turbine diameter, its shape and speed, and the specific gravity of the fluid. Increasing the diameter, speed, or specific gravity requires an increase in the motor size needed to impart the same mixing to the fluid, unless another property is correspondingly reduced. Thus, turbine diameter or speed must be reduced for operation in a fluid whose specific gravity is greater than 1.0 if the motor size is to remain unchanged.

To begin this sizing procedure, the designer should estimate the motor size needed at 0.2 kW/m^3 (1 hp/1,000 gal) of mixing tank capacity and the turbine impeller diameter as one-third the tank diameter. The turbine should be located 1 turbine diameter above the tank bottom. The following procedure is then applied to determine the turbine speed and the corrected turbine diameter. Tables 4-1 and 4-2, relating ideal turbine diameter, speed, and motor horsepower, have been compiled for radial and axial turbines with one and two blades, operating in water. The procedure for axial turbines is the same as

TABLE 4-1

DIAMETER (IN) FOR RADIAL TURBINES IN WATER
(SINGLE AND DUAL TURBINES)

r/min[a]	Motor horsepower																					
	2		3		5		7½		10		15		20		25		30		40		50	
	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT
420	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
350	--	--	--	--	--	--	--	--	--	--	16	--	17	--	18	--	19	--	20	--	21	--
280	--	--	--	--	--	--	--	--	--	--	19	17	20	18	21	19	22	20	23	21	24	22
230	15	13	16	14	18	16	19	17	20	18	21	19	22	20	24	21	25	22	26	23	27	25
190	16	14	17	15	19	17	21	19	22	20	24	22	25	23	27	24	28	25	29	27	31	28
175	17	15	18	16	21	18	23	21	24	22	26	24	27	25	29	26	30	27	31	29	33	30
155	18	16	20	17	22	19	24	22	26	23	28	25	29	27	30	28	32	29	33	30	35	32
125	21	19	23	21	26	23	28	25	29	26	31	28	33	30	34	32	36	33	39	34	40	36
115	23	21	25	23	28	25	30	27	31	28	34	31	36	33	37	34	39	35	42	37	43	39
100	24	22	26	24	29	26	32	28	33	30	36	33	38	35	40	36	41	37	44	40	46	41
84	27	24	29	26	32	29	34	32	36	34	39	36	42	39	44	40	46	41	49	43	51	46
68	30	27	33	30	36	33	38	35	41	38	45	41	48	44	51	46	53	47	54	50	57	52
45	39	35	42	38	46	42	51	46	54	48	58	51	62	54	64	57	68	60	70	62	74	66
37	44	40	47	42	53	47	56	50	60	54	66	59	70	64	72	66	74	68	76	72	82	76
30	50	46	54	48	59	54	66	58	70	62	76	66	80	72	82	74	86	76	90	82	--	--

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[a] Rotational speed

TABLE 4-2

DIAMETER (IN) FOR AXIAL TURBINES IN WATER
(SINGLE AND DUAL TURBINES)

r/min[a]	Motor horsepower																					
	2		3		5		7½		10		15		20		25		30		40		50	
	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT	ST	DT
420	11	10	12	11	14	12	15	13	16	14	17	15	18	16	19	17	20	18	21	19	22	20
350	13	11	14	13	15	14	17	15	18	16	19	17	20	18	21	19	22	19	24	20	26	21
280	15	13	16	14	18	16	19	17	20	18	22	19	24	20	25	21	26	22	27	24	28	25
230	18	15	19	16	20	18	21	19	23	20	25	21	26	23	28	24	29	26	31	26	32	28
190	19	17	20	18	22	19	25	21	26	22	28	25	30	26	31	27	32	28	34	30	36	31
175	20	18	21	19	24	20	26	22	27	24	30	26	32	27	32	28	33	30	36	32	38	32
155	21	19	23	20	26	22	28	25	29	26	32	27	33	29	35	31	37	32	38	33	40	35
125	24	20	26	23	29	26	32	27	33	29	36	32	38	33	39	35	41	36	44	38	45	40
115	26	22	28	24	31	26	32	29	35	31	38	33	40	35	42	37	44	38	47	40	49	42
100	28	25	30	26	33	29	36	32	38	33	41	36	44	38	46	40	48	41	50	44	52	46
84	31	27	33	29	37	32	39	35	42	37	45	40	48	42	50	44	52	45	56	49	58	50
68	35	31	38	33	42	37	44	40	48	42	52	45	55	48	57	50	59	52	64	56	66	58
56	39	35	43	38	48	42	51	44	55	48	59	51	64	55	66	56	68	60	72	64	74	66
45	45	40	49	43	55	48	59	52	62	55	68	60	72	64	74	66	78	68	82	72	86	76
37	51	44	56	49	62	54	66	58	70	62	76	68	80	70	84	74	88	76	--	--	--	--
30	57	50	62	55	68	60	74	66	78	70	86	74	90	80	--	--	--	--	--	--	--	--

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[a] Rotational speed

that for radial turbines.

1. Using the estimated horsepower and turbine diameter values, read the required turbine speed from Table 4-1 or 4-2.

2. Choose all applicable correction factors from Table 4-3 or 4-4, and multiply them to obtain a Power Correction Factor (PCF).

3. Refer to Table 4-4 for the turbine diameter correction needed, depending on the PCF. If the PCF value is between 0.95 and 1.10, the turbine diameter does not need to be corrected. Otherwise, round the turbine diameter to the nearest whole centimeter (inch). If unsure which way to round the number, round down to the lower diameter.

4. If the diameter has been changed, check the new turbine diameter-to-tank diameter ratio. If it is less than the desired value of 0.33, reduce the turbine speed to the next lower value and select the corresponding turbine diameter. Repeat steps 2 through 4 with these new values.

This procedure has been outlined for the designer primarily for awareness of basic factors when purchasing a mixer or mixing tank from a vendor. If the designer, however, chooses to fabricate a tank for a given application, he can exercise this option.

4.6 Flocculant Mixing

Flocculation refers to the physical and chemical process in which suspended solids form larger, faster settling particles. This process immediately precedes sedimentation. Flocculation is accomplished in AMD treatment by the addition of chemical polymers that act as bridging agents between the suspended particles. The polymer solution is prepared in a mix tank and then added to neutralized mine drainage before the clarifier or sedimentation basin. The flocculant reaction is achieved through gentle mixing, which can be done in a vessel equipped with a flocculation chamber. The most common method is by normal turbulence in the transfer pipe to the clarifier. The latter method, however, cannot always be assumed to be adequate or efficient. Many flocculants require a specific reaction time to form the necessary bridging for improved settling. Thus, this common simple method can be wasteful and expensive.

The polymer makeup tank should be stirred by a low-speed, portable propeller mixer (less than 400 r/min) to avoid destroying the polymer molecules. A constant supply of polymer solution is usually fed from a storage tank by a volumetric (positive displacement) pump. Centrifugal pumps should not be used because they can destroy the polymer molecules by shear force. The tank size depends upon the polymer feed rate, the feed concentration, and the solution makeup schedule. Care must be taken to avoid formation of gummy, clogging masses of polymer during the solution makeup.

TABLE 4-3

RADIAL TURBINE PROXIMITY AND LIQUID PROPERTIES FACTORS

Condition Different from Base Case	Factor Single Turbine	 Dual Turbine
C/D = 0.4-0.7	0.95	1.0
C/D = Less than 0.4	0.85	0.92
In unbaffled square or rectangular tank	0.85	0.85
Specific gravity other than 1.0	sp gr	sp gr
Viscosity 1-10,000 cP (centipoise)	1.0	1.0

TABLE 4-4

AXIAL TURBINE PROXIMITY AND LIQUID PROPERTIES FACTORS

Condition Different From Base Case		Factor Single Turbine	 Dual Turbine
C/D = 0.6-0.9		1.05	1.0
C/D = 0.3-0.6		1.10	1.05
C/D = Less than 0.3		Not used	Not used
Cover (Z-C) = 0.5D to 1.0D		--	0.95
In unbaffled square or rectangular tank		0.85	0.85
Specific gravity other than 1.0		sp gr	sp gr
Viscosity, cP	200	1.0	1.0
	500	1.1	1.1
	1,000	1.2	1.2
	2,500	1.3	1.3
	5,000	1.5	1.5
	10,000	1.8	1.8

Rapid mixing of the polymer solution with mine drainage flow is desired to insure adequate polymer dispersion. A small flash mix tank providing a detention time of 10-30 seconds, with a mixer providing 26-53 kW/1,000 l/s (2.2-4.4 hp/1,000 gal/min) of flow, can be used for this purpose, as well as in-line mixers (4). Common practice is to inject polymer into a pipe or flow-splitter box where the turbulence provides mixing. This is acceptable as long as the turbulence and residence time provides complete mixing.

Gentle agitation is desired to promote floc growth. Mechanical flocculators equipped with slowly rotating paddles (less than 60 r/min) serve this purpose. Standard designs are available that can provide 30-45 minutes for floc formation (4). Again, these units have not found much use in AMD treatment plants, because designers have relied upon process turbulence to enhance floc formation. Flocculator-clarifiers combine the two unit processes, eliminating the need for a separate flocculation unit; however, these units are rarely used.

4.7 Summary

There are basically three applications of mixing in AMD treatment. Mixers are used in lime slakers and lime slurry storage tanks, flash mix tanks, and flocculation tanks. The recommended designs for each of these are summarized below.

Slurry mix tanks exist in a wide range of sizes. Smaller tanks, less than 11.355 m^3 (3,000 gal), are equipped with either angular-mounted, portable propeller mixers or with fixed-mounted propeller mixers with baffles. Large tanks, greater than 11.355 m^3 (3,000 gal), are usually equipped with top-mounted turbine mixers and also contain baffles. Mixer horsepower requirements can be estimated from the general rule of 0.2 kW/m^3 (1 hp/1,000 gal).

Flash mix tanks are sized to provide a detention time between 10 and 30 seconds. The mixer employed is almost always an angular-mounted, portable propeller mixer sized accordingly, but limited to 2.24 kW (3.0 hp).

To enhance chemical coagulation and improve settling performance, mechanical flocculators can be used. A flocculation detention time of 3-5 minutes should be provided before settling. Several types of flocculators are available, including vertical and horizontal paddle and turbine units. Flocculator suppliers should be contacted to provide detailed information on the basin and flocculation unit.

In conclusion, this chapter has presented the more basic, common sense factors influencing mixer and mixing tank design. The designer should be aware that good and complete mixing is not always easily accomplished.

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CHAPTER 5

IRON OXIDATION

5.1 Introduction

Acid mine drainage often contains significant concentrations of iron resulting from the oxidation of pyritic minerals associated with coal. The mining process exposes these pyrites or iron sulfides to the atmosphere and to moisture, thus causing their oxidation to soluble ferrous sulfate salt and sulfuric acid. These salts readily dissolve in water forming mine drainage. If there is an overabundance of acid salts to the alkalinity in the water, the mine drainage will be acidic.

As mine drainage is formed, iron is first present in the ferrous (Fe^{++}) form. Ferric iron is much less soluble than ferrous and can be precipitated as a hydroxide to effluent quality levels below the minimum allowable pH of 6.0 (1, 2). As shown in Figure 5-1, minimum ferric solubility occurs at a pH of 8.0, while ferrous does not reach minimum solubility until the pH approaches 11.0 (2, 3). At the maximum allowable discharge pH of 9.0, ferrous iron is soluble to about 4 mg/l, which is in excess of the discharge limitations for new sources. Therefore, in most mine drainage treatment systems, such as any of the chemical neutralization processes, it is imperative to oxidize any ferrous iron to the ferric form so it can be effectively removed at a lower system pH. The methods available to accomplish this oxidation are natural or mechanical aeration, chemical oxidation, and biological systems.

5.2 Aeration Systems

Ferrous iron, when exposed to oxygen, will oxidize to ferric iron at a rate determined by the ferrous iron concentration, the dissolved oxygen concentration, and the pH of the solution. At pH values greater than 6.0, the reaction occurs according to the following rate equation:

$$\frac{-d\ (Fe^{++})}{dt} = k\ (Fe^{++})\ (O_2)\ (OH^-)^2 \tag{9}$$

The reaction is first order with respect to the ferrous iron and the dissolved oxygen concentrations. This means that the oxidation rate decreases as the concentration of either decreases. The reaction rate is second order with respect to the hydroxyl ion (OH^-) concentration for pH values greater than 6.0. Thus, the reaction rate increases 100 times for each one-unit rise

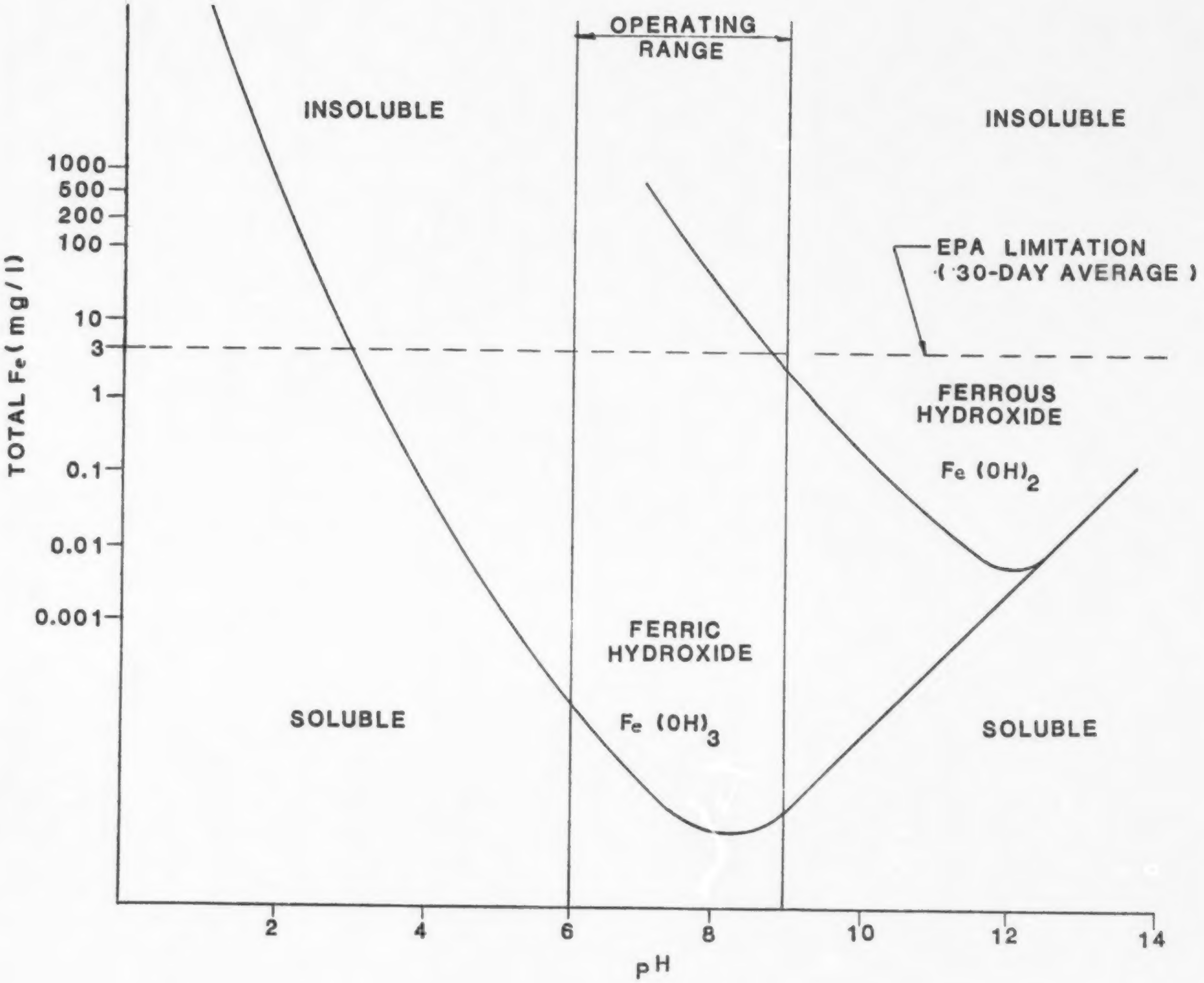

Figure 5-1. Solubility of ferric and ferrous iron at various pH.

in pH above pH 6.0. The rate of ferrous iron oxidation can be classified as extremely slow (days) at a pH of less than 3.0, slow in the pH 3.0-6.0 range, moderate to fast in the pH 6.0-8.0 range, and rapid above this point (3).

Ferric iron will precipitate as ferric hydroxide sufficiently to meet effluent limits at a pH of 5.0. At this pH, unfortunately, the oxidation rate for ferrous iron is slow. The reaction does not increase to an acceptable rate (minutes) until the pH is 8.0 or greater. When iron is mostly in the ferrous form, aeration processes are most efficiently operated within a pH range of 8.0-9.0. In this pH range, the oxidation reaction takes place in a matter of minutes, and the controlling parameter for the design of the aeration unit becomes a function of the oxygen transfer efficiency and not the chemical reaction of oxygen and iron. The aerator should be sized to provide dissolved oxygen saturation in the aeration basin, assuming maximum oxygen transfer.

5.2.1 Oxygen Requirements

The amount of iron to be oxidized in a given time will determine the capacity of the aeration system. If the system cannot meet this oxygen requirement, then oxidation will be incomplete. The rate of oxidation increases as the dissolved oxygen concentration in water increases to its saturation point. Any additional aeration capacity beyond saturation does not benefit the ferrous iron oxidation rate.

The rate of oxygen transfer into water depends on the initial oxygen deficit; i.e., the lower the initial dissolved oxygen concentration, the easier it is to dissolve oxygen by aeration. Conversely, the ferrous iron oxidation rate depends on the dissolved oxygen concentration, with the maximum rate occurring at saturation. A compromise between these two mechanisms must be met to optimize the aeration process. In water with a pH greater than 8.0, the oxidation rate is rapid enough that maintaining dissolved oxygen near saturation is unnecessary. If aeration is performed at a pH less than 8.0, then a fairly high dissolved oxygen level must be maintained for efficient ferrous iron oxidation (4).

The chemical equations for the oxidation of ferrous iron to ferric and its subsequent hydrolysis are as follows:

$$Fe^{+2} + \tfrac{1}{4}O_2 + H^+ \rightarrow Fe^{+3} + \tfrac{1}{2}H_2O \qquad (10)$$

$$Fe^{+3} + 3H_2O \rightarrow Fe(OH)_3 + 3H^+ \qquad (11)$$

The stoichiometric relationship of this equation indicates that 1 kg of oxygen will oxidize 7 kg of ferrous iron under ideal conditions. During this oxidation and hydrolysis reaction, 1 mol of acidity (as H_2SO_4) is formed for each mole of ferrous iron that is oxidized. Therefore, sufficient excess alkalinity must be added during neutralization to compensate for the acidity formed and to maintain optimum pH conditions during aeration for ferrous iron oxidation.

Whatever type aeration system is used, it must meet the oxygen demand for ferrous iron oxidation. The theoretical oxygen demand for any mine water can be calculated by Equation 12 (4).

$$O_2 = Qw \times Fe \times 5.16 \times 10^{-4} \tag{12}$$

where O_2 = Theoretical O_2 demand (kg O_2/hr)

Qw = Acid mine drainage flow rate (l/s)

Fe = Fe^{+2} initial concentration (mg/l)

$$O_2 = Qw \times Fe \times 7.14 \times 10^{-5}$$

where O_2 = Theoretical O_2 demand (lb O_2/hr)

Qw = Acid mine drainage flow rate (gal/min)

Fe = Fe^{+2} initial concentration (mg/l)

Oxygen makes up about 21% of air by volume. Only a fraction of the oxygen that comes in contact with the water is actually absorbed. This fraction, expressed as the oxygen transfer efficiency, differs for each aeration system under actual operation conditions. The total air needed to supply the theoretical oxygen demand, taking into account the above considerations, can be calculated by Equation 13 (4).

$$Qa = \frac{6.324 \times O_2}{E} \tag{13}$$

where Qa = Total air demand (standard m^3/min)

O_2 = Theoretical oxygen demand (kg/hr of oxygen required)

E = Oxygen transfer efficiency (as %)

$$Qa = \frac{101.36 \times O_2}{E}$$

where Qa = Total air demand (standard ft^3/min)

O = Theoretical oxygen demand (lb/hr of oxygen required)

E = Oxygen transfer efficiency (as %)

Oxygen transfer efficiencies (E) range from 3% to 25%, depending upon the type and size of aerator used and the depth of submergence. Manufacturers of aeration equipment should be consulted for the transfer efficiency of their specific aeration system. A 10% transfer efficiency is generally used when the exact value is not available. The maximum air requirements will be at the highest anticipated air temperature when the density is least.

In addition to providing the required oxygen for ferrous iron oxidation, the aeration system must be capable of keeping the ferric hydroxide solids and unreacted reagent in suspension. If there is not sufficient mixing, these solids will settle to the bottom of the aeration basin. Settled solids, in effect, reduce the aeration volume and aeration time, leading to incomplete ferrous iron oxidation. The aerator must be sized to meet both oxygen and mixing requirements. In mine drainage with very high iron concentrations, the power required for oxygen transfer is usually sufficient to meet the mixing requirements. In most other cases, the aerator size may be determined by the mixing requirements.

Aeration is the conventional method used for iron oxidation. The aeration processes, which solubilize atmospheric oxygen in mine drainage, can be classified into four categories; i.e., mechanical surface aeration, diffused air aeration, submerged turbine aeration, and cascade aeration.

5.2.1.1 Mechanical Surface Aeration

Mechanical surface aeration introduces atmospheric oxygen into water by rotating blades positioned below the static water level in an aeration basin. The turbulence created by the aerator disperses air bubbles and keeps the iron floc in suspension. Oxygen is absorbed by the water at the air-water interface following the classical laws of gas absorption. The dissolved oxygen then reacts with the ferrous iron to complete the reaction.

Mechanical surface aerators (Figure 5-2) are the more popular choice by the designers of mine drainage treatment plant aeration systems. They can be either slow-speed turbines or high-shear, high-speed, axial flow turbines. Both types pull water from beneath the rotor blades and spray it across the water surface. The oxygen transfer occurs during this splashing. Mechanical aerators are either floating or structurally supported. The oxygen transfer efficiency of surface aerators is generally the highest of the aeration systems considered, usually in the 1.8- to 2.1-kg O_2/kW-hr (3.0-3.5 lb O_2/hp-hr) range (5). Floating aerators perform best in circular basins. These basins should be limited to a depth of 3.0 m (10 ft) to eliminate the need for a draft tube or submerged piping. Slow-speed aerators are generally used because more water is pumped and a better efficiency is obtained for the same horsepower than with a high-speed aerator. Mechanical aerators can be used in both square and circular basin configurations.

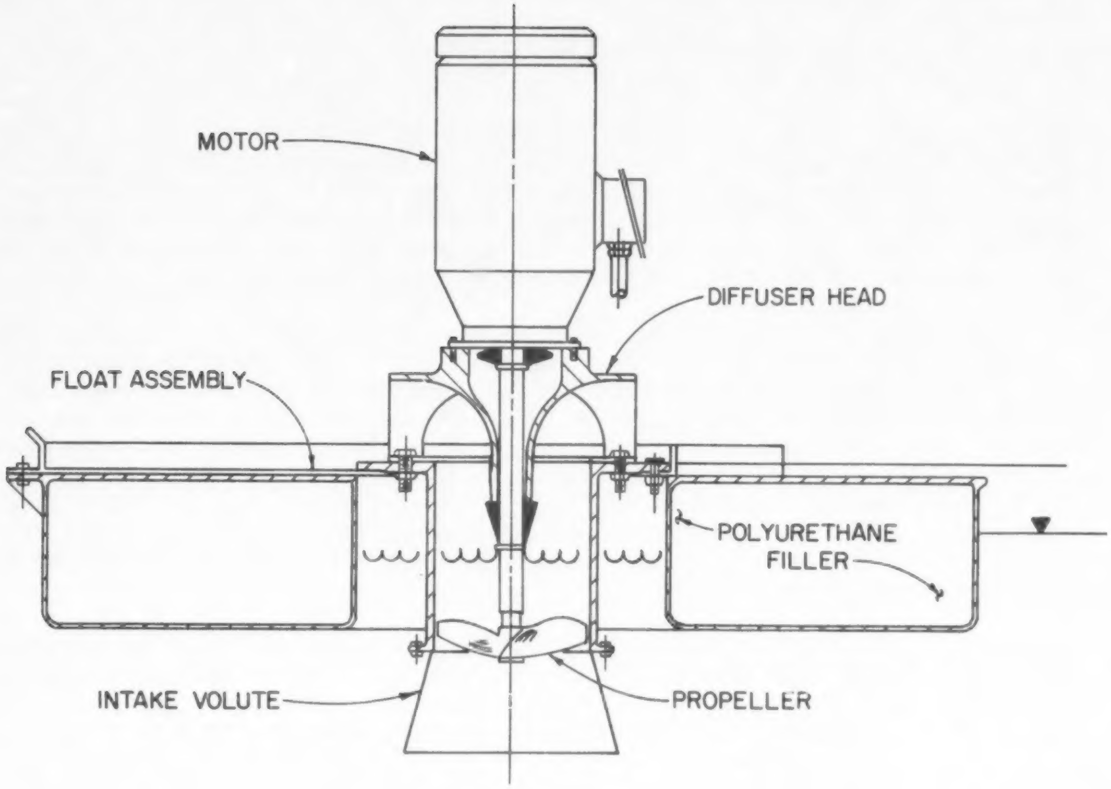

Figure 5-2. Typical floating mechanical aerator.

5.2.1.2 Diffused Air Aeration

Diffused air aeration systems have rarely been used for the treatment of acid mine drainage. Although diffused systems are used extensively in sewage treatment, gypsum and iron or aluminum precipitates encountered with AMD tend to make this system unsuitable for this application.

Diffused air systems introduce air, supplied by a blower, to the water through diffusers placed near the tank bottom. Oxygen transfer occurs as the air bubbles rise to the surface. The diffuser devices deliver air through a porous medium, such as carborundum, nylon, or saran. These have a better oxygen transfer efficiency than coarse bubble diffusers, which deliver air through perforated pipes. This advantage is offset, though, by the cleaning and replacement costs caused by clogging of the porous medium with precipitates in the water. A gypsum problem resulting from the treatment of AMD would prohibit the use of a diffused air system. Coarse bubble devices generally have a lower oxygen transfer efficiency and a higher capital cost than mechanical aerators, making them impractical for AMD treatment.

The air and oxygen requirements for diffused aeration have already been specified (see Equations 12 and 13). The manufacturer will specify the optimum configuration and spacing for the desired oxygen transfer rate and mixing needs. Diffused air aeration basins can be constructed of concrete, are usually rectangular in shape, and operate at depths of 3.7-4.6 m (12-15 ft). The diffusers work most efficiently when placed to one side in a single line along the horizontal axis of the basin (4). The basin design and diffuser placement are critical for good mixing and oxygen transfer.

5.2.1.3 Submerged Turbine Aeration

Submerged turbine aerators combine both of the previous methods. An air sparger is located near the bottom of the tank and above it is one or more turbine impellers. The shearing action of the rotating impeller produces small bubbles necessary for good oxygen transfer efficiency.

Submerged turbine aerators offer few benefits over mechanical surface aerators. The higher degree of oxygen transfer is of little benefit to ferrous iron oxidation. These are less efficient than surface aerators and require more power to meet the oxygen demand. Standard oxygen transfer efficiencies are 1.0-1.3 kg O_2/kW-hr (1.5-2.0 lb O_2/hp-hr) for single impeller turbines, and 1.7-2.1 kg O_2/kW-hr (2.5-3.0 lb O_2/hp-hr) for dual impeller turbines (6). Submerged piping is also required for these turbine units. The submerged turbine units are fixed-mounted and operate well in winter conditions.

5.2.1.4 Cascade Aeration

Aeration by gravity or cascading is a practical, inexpensive, and popular method employed for oxidizing low ferrous iron concentrations in mine drainage (less than 50 mg/l). The most common type of cascade aerator is an open trough, usually a half-round pipe, lined with splash blocks to induce turbulence and increase aeration.

Other methods of cascade aeration are stairsteps, falls, a combination of both, and wide, shallow, open flumes.

Assigning a scientific or technical approach to the design of any type of cascade aerator is difficult and, for the most part, can be classified as a rough estimate. Many variables, such as mine water temperature, initial oxygen deficit, oxygen transfer, and air temperature, influence aeration efficiency. It is difficult to predict efficiency unless actually performed at full scale.

Therefore, this chapter will present an estimate of the design of an aeration trough. The designer can then refine the device (i.e., increase length of the trough, insert more splash blocks) when the unit goes on-line. Also, this uncertainty will require the designer to provide flexibility and margin for error in the design and erection.

5.2.2 Aeration Trough Design

The maximum rate of iron oxidation occurs at pH 8.5 provided oxygen is available. Therefore, an operating pH less than ideal will require longer aeration times and usually renders this method impractical for waters with high ferrous iron concentrations (greater than 100 mg/l).

Detention times for this method are generally less than 5 minutes. The detention time for any trough can be found using Equations 14 and 15.

$$V = A\ (L) \tag{14}$$

where V = volume of trough m^3 (ft^3)

A = cross-sectional area of flow m^2 (ft^2)

L = length of trough m (ft)

$$D_t = \frac{V}{Q} \tag{15}$$

where D_t = detention time (min)

V = volume of trough m^3 (ft^3)

Q = m^3/min (gal/min)

The difficult part of the equation is computing the cross-sectional area of the flow in the pipe at a given depth. To alleviate this problem, Table 5-1 has been provided. This table presents the cross-sectional area per linear meter (foot) of the most commonly used pipes, 15.2-121.9 cm (6-48 in), flowing at various depths. Only the lower depths, those less than half the diameter, are applicable. Volume is merely the cross-sectional area multiplied by the trough length (see Equation 14).

With the volume of the trough in cubic meters (gallons) known, the detention time can be easily computed. A rule of thumb among designers has been 0.1 min/mg/l of ferrous iron at pH 8.5. Therefore, if a mine drainage contains 30 mg/l of ferrous iron neutralized to 8.5, 3 minutes of trough detention time with vigorous aeration should be provided. The factor is a good starting point but in no way guarantees complete oxidation. This can be determined only in the field by analysis of the drainage.

The number of splash blocks, size, and spacing are chosen arbitrarily.

TABLE 5-1

CROSS-SECTIONAL AREA OF FLOW IN CIRCULAR PIPE[a]

Depth of Flow (d)		Diameter of Pipe (D)															
cm	in	cm 15.2	in 6.0	cm 20.3	in 8.0	cm 25.4	in 10.0	cm 30.5	in 12.0	cm 38.1	in 15.0	cm 45.7	in 18.0	cm 53.3	in 21.0	cm 61.0	in 24.0
1.27	0.50	.001	.008	.001	.009	.001	.010	.001	.011	.001	.013	.001	.014	.001	.015	.001	.016
1.91	0.75	.001	.014	.002	.017	.002	.019	.002	.020	.002	.023	--	--	--	--	--	--
2.54	1.00	.002	.022	.002	.025	.003	.028	.003	.031	.003	.035	.004	.039	.004	.042	.004	.045
3.18	1.25	.003	.030	.003	.035	.004	.039	.004	.043	.005	.049	--	--	--	--	--	--
3.81	1.50	.004	.038	.004	.045	.005	.051	.005	.057	.006	.064	.007	.070	.007	.076	.008	.082
4.45	1.75	.004	.048	.005	.056	.006	.064	.007	.071	.007	.080	--	--	--	--	--	--
5.08	2.00	.005	.057	.006	.068	.007	.078	.008	.086	.009	.097	.010	.107	.011	.117	.012	.125
5.72	2.25	--	--	--	--	.009	.092	.009	.102	.011	.115	--	--	--	--	--	--
6.35	2.50	--	--	--	--	.010	.107	.011	.119	.012	.134	.014	.149	.015	.162	.016	.174
6.99	2.75	--	--	--	--	--	--	.013	.136	.014	.154	--	--	--	--	--	--
7.62	3.00	--	--	--	--	--	--	.014	.154	.016	.175	.018	.197	.020	.211	.021	.227
8.26	3.25	--	--	--	--	--	--	--	--	.018	.196	--	--	--	--	--	--
8.89	3.50	--	--	--	--	--	--	--	--	.020	.218	.022	.242	.024	.264	.026	.284
9.53	3.75	--	--	--	--	--	--	--	--	.022	.240	--	--	--	--	--	--
10.16	4.00	--	--	--	--	--	--	--	--	--	--	.027	.292	.030	.319	.032	.344
10.80	4.25	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
11.43	4.50	--	--	--	--	--	--	--	--	--	--	.032	.345	.035	.378	.038	.408
12.07	4.75	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
12.70	5.00	--	--	--	--	--	--	--	--	--	--	--	--	.041	.439	.044	.474
13.34	5.25	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--	--
13.97	5.50	--	--	--	--	--	--	--	--	--	--	--	--	.047	.502	.050	.543

---

[a] The units for the area of flow in pipe are m^2 when the diameter is given in cm and ft^2 when the diameter is given in in.

TABLE 5-1 (continued)

Depth of Flow (d)		Diameter of Pipe (D)													
cm	in	cm 58.6	in 27.0	cm 76.2	in 30.0	cm 83.8	in 33.0	cm 91.9	in 36.0	cm 121.9	in 48.0	cm 137.2	in 54.0	cm 152.4	in 60.0
1.27	0.5	.002	.017	.002	.018	.002	.019	.002	.020	.002	.023	.002	.024	.002	.025
2.54	1.0	.004	.048	.005	.050	.005	.053	.005	.055	.006	.064	.006	.064	.007	.071
3.81	1.5	.008	.087	.009	.092	.009	.096	.009	.101	.011	.110	.012	.124	.012	.131
5.08	2.0	.012	.133	.013	.141	.014	.148	.014	.154	.017	.179	.018	.190	.019	.201
6.35	2.5	.017	.185	.018	.195	.019	.205	.020	.214	.023	.250	.025	.265	.026	.280
7.62	3.0	.022	.241	.024	.255	.025	.269	.026	.281	.030	.327	.032	.348	.034	.367
8.89	3.5	.028	.302	.030	.320	.031	.337	.033	.353	.038	.411	.041	.437	.043	.461
10.16	4.0	.034	.367	.036	.389	.038	.410	.040	.429	.046	.500	.049	.532	.052	.562
11.43	4.5	.040	.436	.043	.462	.045	.486	.047	.510	.055	.595	.059	.633	.062	.669
12.70	5.0	.047	.507	.050	.538	.053	.567	.055	.595	.065	.694	.069	.739	.073	.782
13.97	5.5	.054	.581	.057	.617	.060	.651	.063	.683	.074	.798	.079	.850	.084	.899
15.24	6.0	.061	.658	.065	.699	.069	.738	.072	.774	.084	.907	.090	.966	.095	1.022
16.51	6.5	.068	.737	.073	.783	.077	.827	.081	.869	.095	1.019	.101	1.086	.107	1.149
17.78	7.0	--	--	.081	.870	.085	.920	.090	.967	.105	1.135	.112	1.210	.119	1.281
19.05	7.5	--	--	.089	.960	.094	1.015	.099	1.067	.116	1.254	.124	1.338	.132	1.417
20.32	8.0	--	--	--	--	.103	1.112	.109	1.170	.128	1.377	.136	1.469	.145	1.556
21.59	8.5	--	--	--	--	.113	1.211	.118	1.275	.140	1.502	.149	1.604	.158	1.700
22.86	9.0	--	--	--	--	--	--	.138	1.382	.152	1.631	.162	1.742	.172	1.847
24.13	9.5	--	--	--	--	--	--	--	--	.164	1.763	.175	1.884	.186	1.997
25.40	10.0	--	--	--	--	--	--	--	--	.176	1.897	.188	2.028	.200	2.151
26.67	10.5	--	--	--	--	--	--	--	--	.189	2.033	.202	2.175	.214	2.308
27.94	11.0	--	--	--	--	--	--	--	--	.202	2.172	.216	2.325	.229	2.468
29.21	11.5	--	--	--	--	--	--	--	--	.215	2.313	.230	2.477	.244	2.630
30.48	12.0	--	--	--	--	--	--	--	--	.228	2.457	.244	2.632	.260	2.796
31.75	12.5	--	--	--	--	--	--	--	--	--	--	.259	2.789	.275	2.964
33.02	13.0	--	--	--	--	--	--	--	--	--	--	.274	2.948	.291	3.134
34.29	13.5	--	--	--	--	--	--	--	--	--	--	.289	3.109	.307	3.307
35.56	14.0	--	--	--	--	--	--	--	--	--	--	--	--	.323	3.482
36.83	14.5	--	--	--	--	--	--	--	--	--	--	--	--	.340	3.659
38.10	15.0	--	--	--	--	--	--	--	--	--	--	--	--	.357	3.839

Splash blocks are usually tack-welded to the bottom of the trough (corrugated metal pipe) and inclined to cast the flow into the air. Figure 5-3 shows a typical splash block and the placement pattern. This is one example of many configurations used today. However, any concept for inducing the required turbulence is suitable; the possibilities are limited only by the imagination.

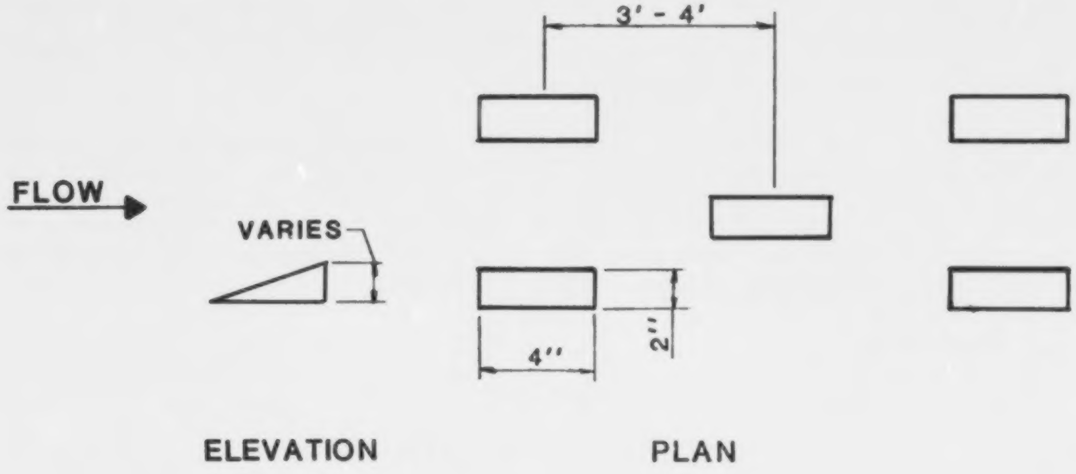

Figure 5-3. Typical splash block placement pattern.

The application of gravitational or cascade aeration is limited by the initial ferrous iron concentration (less than 50 mg/l) and the neutralization pH (8.5). Any variance from these criteria, especially pH, will affect and inhibit complete iron oxidation by this method. A good trough velocity is 0.9-1.2 m/s (3-4 ft/s); flow depth should not exceed one-fourth of the pipe diameter. This will allow adequate freeboard to contain splash and reduce overflow.

When the trough exceeds 366 m or 1.2 m/s x 300 s (1,200 ft or 4 ft/s x 300 s) or when ferrous iron concentrations are greater than 50 mg/l, a mechanical aerator should be considered.

5.2.3 Aeration Basin Design

Proper design of the aeration basin is necessary for the efficient oxidation of ferrous iron. The sizing of this unit is important to insure that adequate, but not excessive, aeration periods are provided. Attention must also be given to the basin plan, depth dimensions, and inlet and outlet structures. Aeration basins are most often earthen units lined with riprap or asphalt, but can also be constructed of concrete or steel.

Mathematical models have been used to predict aeration times for the oxidation of ferrous iron at varying pH ranges (1, 2, 4, 15). Due to the variable nature of mine drainage and the effects other dissolved ions may impose on the reaction, laboratory tests are the most reliable way to optimize the aeration system design. The laboratory tests should follow the method outlined at the end of this chapter. To insure plant operation at peak conditions, the tests should be performed on a sample containing the maximum ferrous iron concentration and at the lowest anticipated operating pH. Also, the tests should be carried out at the desired operating pH or at several others if a comparison of results is desired.

The detention time needed for ferrous iron oxidation, as determined by the laboratory tests, must be multiplied by a safety factor for the design of the full-sized aeration basin to reduce the possibility of short-circuiting through the aeration basin. The chances of short-circuiting are greater with the shorter detention periods. The calculated detention time should be multiplied by the safety factors listed in Table 5-2.

TABLE 5-2

AERATION DETENTION TIME SAFETY FACTORS

Calculated Detention Time min	Safety Factor min
≥ 16	2.0
11-15	3.0
6-10	4.0
3-5	8.0
1-2	10.0
< 1	≥ 10.0

The volume of the aeration basins is easily determined using the mine drainage flow rate and the calculated detention time as follows:

$$V = QD_tf \tag{16}$$

where V = Volume, m^3 (gal)

Q = Flow, m^3/s (gal/min)

D_t = Detention Time, s (min)

f = Safety Factor (from Table 5-2)

The design of the aeration basin must be made in conjunction with the selected aerator. The aerator must be positioned in the basin so the entire volume is aerated and well mixed, thus eliminating dead spots where solids will accumulate. Aerator manufacturers will recommend the surface area and depth for complete oxygen dispersion and mixing. Therefore, the designer must know these limitations. Manufacturers will also specify the aerator size in kilowatts, kW (horsepower, hp), and the oxygen transfer rate, kg O_2/kW-hr (lb O_2/hp-hr). Using this data, an efficient and proper aeration system can be designed.

The oxygen transferred can be calculated by multiplying the aerator kilowattage (horsepower) by the oxygen transfer rate.

$$O_2 \text{ supplied} = O_2 \text{ transfer rate} \times \text{aerator power}$$

$$\text{kg } O_2/\text{hr (lb } O_2/\text{hr)} = \text{kg } O_2/\text{kW-hr} \times \text{kW (lb } O_2/\text{hp-hr} \times \text{hp)} \qquad (17)$$

The designer must determine whether one or more aerators will meet the oxygen requirement. This decision will be based on an evaluation of the capital and operating costs and the ability of the aerator(s) to keep the basin contents mixed.

The size of tank matters also. A floating aerator will generally be effective only in a 7.6-m (25-ft) diameter pattern and a 3.1- to 3.7-m (10- to 12-ft) depth. Typical aeration basin depths range between 1.5 and 4.6 m (5 and 15 ft). The depth selected will depend upon the detention required, the basin surface area, and the mixing characteristics of the aerator(s). In the case of mechanical surface aerators, draft tubes can be added if the basin depth becomes excessive. The draft tube is an extension that causes a deeper intake, increasing the effective operating depth. Turbine and diffused aeration systems have optimum operating depths, and the equipment manufacturer should be consulted.

When a single surface aerator is used, the ideal basin shape is circular. This usually eliminates dead spots where solids accumulate. Also, this geometric shape reduces the possibility of short-circuiting, which reduces aeration time. When using more than one aerator, a rectangular shape may be preferred. For smaller flows, minimum basin sizing as stipulated by the manufacturer may overrule earlier decisions on the basin size as calculated for detention requirements.

Design and construction of the aeration tank should follow standard engineering practice. Steel and concrete construction should follow American Institute of Steel Construction and American Concrete Institute specifications. If an earthen basin is to be used, consideration must be given to the soil and geology, basin location, and construction procedure. Stable slopes should be established, with typical values as 3:1 for the outside slope and 2:1 for the inside slope. The soil should be compacted properly to insure stability. The basin bottom and inside slopes should be lined with clay, concrete, asphalt, or a synthetic liner. When using clay as a liner, riprap at the water periphery will prevent erosion caused by wave action.

Regardless of the type of aeration basin, the inlet and outlet structures must be designed to minimize short-circuiting. They should be located at opposite ends of the aeration tank, allowing maximum use of the aeration tank volume. The water level in the tank is determined by the elevation of the outlet structure. Outlet structures are typically weirs, weir boxes, or open vertical pipes. A baffle should be placed in front of the outlet structure to separate the turbulent aeration zone from the outlet zone. The inlet structure should allow the water to free-fall into the tank, thus eliminating hydraulic problems. A direct discharge by means of a pipe into the aeration tank is adequate, because the mixing from the aerator will disperse the influent water. Provisions for access to both floating and supported aerators should also be considered.

5.3 Chemical Oxidation

Besides aeration, several chemicals have been utilized for iron oxidation. This chapter presents those chemicals that have potential use for low ferrous iron drainages.

5.3.1 Iron Oxidation by Ozone

Ozone (O_3) is a reactive, gaseous allotrope of oxygen. The only practical method presently available to produce ozone is by electric discharge in which a current is passed through a stream of air or oxygen. In 1970, the EPA sponsored a study of the cost-effectiveness of on-site ozone generation facilities and larger central plants that would supply ozone via a distribution system (6). It was concluded that the central plant system was more economical than individual on-site ozone generators except for low-iron (less than 50 mg/l), low-flow (less than 946 m^3/d) conditions.

Theoretically, 1.0 kg (2.2 lb) of ozone will oxidize 2.3 kg (5.1 lb) of ferrous iron. The same amount of acid is released during ozone oxidation as during aeration. Assuming 86% ozone utilization, 1.0 kg (2.2 lb) of ozone will oxidize 2.0 kg (4.4 lb) of ferrous iron (6).

The benefits of ozone treatment over lime aeration are as follows: (1) the oxidation reaction is efficient and quick, allowing use of smaller reaction basin; (2) close process control needed for lime treatment is not needed for ozone treatment; and (3) neutralization to pH 6.0 is all that is required, allowing treatment with limestone as well as other alkalis. The sludge produced by limestone-ozone combination is denser than lime sludge, reducing sludge handling requirements.

5.3.2 Iron Oxidation by Hydrogen Peroxide

The oxidation of iron using hydrogen peroxide (H_2O_2) merits consideration where "specialty" conditions exist. These conditions can be generally classified as follows:

1. alkaline mine drainages (pH greater than 6.0) with low oxygen requirements for iron oxidation;
2. where pH adjustment is made merely for more favorable iron oxidation rates;
3. as a supplemental source where existing facilities become iron-overloaded and expansion is impossible.

No steadfast rules can be presented as to when to consider hydrogen peroxide. An overall economic evaluation of the particular application must be performed. Comparing reagent cost against mechanical aerator operating cost, which includes capital reinvestment for equipment and amortization, indicates that the hydrogen peroxide oxidation method was more expensive; however, a blanket elimination should not be made for every treatment plant.

The general chemical characteristics of this reagent follow, along with the simple stoichiometric relationship necessary for estimating tne theoretical requirement.

5.3.2.1 Physical Properties, Handling, and Storage of Hydrogen Peroxide

Hydrogen peroxide is colorless, has a distinctive pungent odor, and is completely miscible with water. It is prepared in three commercial grades, 35%, 50%, and 70% by weight; the remaining percentage is water (see Table 5-3) (7, 8). Hydrogen peroxide is neither poisonous nor flammable, but it strongly irritates skin, mucous membrane, and eye tissue; it also decomposes to form oxygen, which supports combustion. Additives keep decomposition of uncontaminated solutions to about 2%/year. Many metals, salts, dust, dirt, oil, and rust greatly increase the decomposition rate. Thus, proper handling and storage procedures must be followed.

Standard means of shipment of hydrogen peroxide include 57-1 (15-gal), 114-1 (30-gal), and 208-1 (55-gal) polyethylene-lined drums; 114-1 (30-gal) aluminum drums; special tank trucks, 7,570-15,140 1 (2,000-4,000 gal); and tank cars, 15,140, 22,710, and 30,280 1 (4,000, 6,000, and 8,000 gal). The drums should be accompanied with special drum rockers, wrenches, and spouts.

Hydrogen peroxide should be stored only in the original containers or in

TABLE 5-3

PHYSICAL PROPERTIES OF HYDROGEN PEROXIDE

H_2O_2 concentrations, weight %	35	50	70	90	98
Active oxygen content, weight %	16.5	23.5	32.9	42.3	46.1
g H_2O_2/l	396	600	903	1248	1407
Specific gravity 20°C/4°C	1.133	1.20	1.29	1.387	1.436
lb/gal at 20°C	9.4	10.0	10.8	11.6	11.95
Boiling point, °C	108	114	126	141	149
Boiling point, °F	226	237	258	286	300
Freezing point, °C	-33	-52	-40	-11	-3
Freezing point, °F	-27	-62	-40	12	28
Viscosity 20°C (cP)	1.11	1.17	1.24	1.26	1.25
Total vapor pressure at 30°C (mm Hg)	23.33	18.3	10.1	5	3
Heat of dilution cal/g mol of H_2O_2 at 25°C and 1 atm	-84	-178	-381	--	--

properly designed tanks made of properly prepared compatible material. Storage tanks for hydrogen peroxide must be vented and stored in a clean, fireproof area. Hydrogen peroxide suppliers should be consulted for detailed design requirements (see Figure 5-4).

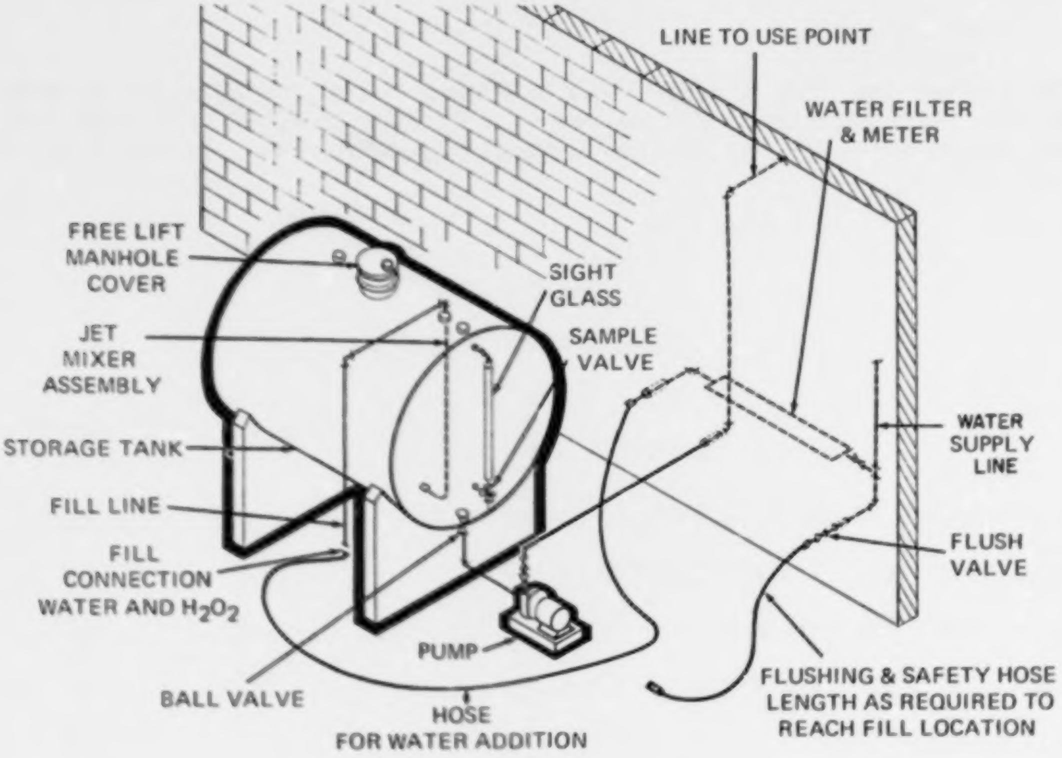
LINE TO USE POINT
WATER FILTER & METER
FREE LIFT MANHOLE COVER
SIGHT GLASS
JET MIXER ASSEMBLY
SAMPLE VALVE
STORAGE TANK
WATER SUPPLY LINE
FILL LINE
FILL CONNECTION WATER AND H_2O_2
FLUSH VALVE
PUMP
BALL VALVE
HOSE FOR WATER ADDITION
FLUSHING & SAFETY HOSE LENGTH AS REQUIRED TO REACH FILL LOCATION

Figure 5-4. Hydrogen peroxide feeding system.

5.3.2.2 Chemistry

Equation 18 expresses the reaction of ferrous iron with hydrogen peroxide (H_2O_2).

$$H_2O_2 + 2Fe^{++} + 2H^{+} \rightarrow 2Fe^{+++} + 2H_2O \qquad (18)$$

One mole of hydrogen peroxide oxidizes 2 mol of iron. Expressed as a weight ratio, 0.45 kg (1.0 lb) of H_2O_2 will oxidize 1.5 kg (3.3 lb) of ferrous iron. The oxidized iron is then available to react with alkalinity to form insoluble ferric hydroxide.

Using Equation 18 and knowing the ferrous iron loading, the theoretical hydrogen peroxide requirement can easily be calculated.

5.3.2.3 Cost

The capital cost for a portable or permanent feeder installation is about $5,000. This includes duplicate pumps, a 19,000-l (5,000-gal) storage tank, and foundation. The reagent costs freight on board (FOB) are presented in Table 5-4.

TABLE 5-4

HYDROGEN PEROXIDE COSTS

	¢/kg	(¢/lb)
Bulk delivery 17,000 l (4,500 gal) tanker @ 50%	9.7	21.5
Bulk, 1,900-2,300 l small (500-600 gal) @ 50%	10.9	24
Drums, 208 l (55 gal) @ 50%	12.25	27
Drums, 114 l (30 gal) @ 50%	13.64	30
Full strength (100%)	19.5	43

The daily reagent cost can be determined based on the desired method of delivery.

5.3.3 Summary

The oxidation of iron with hydrogen peroxide can be more convenient than economical, but merits consideration for certain applications. It can be the primary source of oxygen, but is best used in conjunction with existing aeration facilities. Proved feasible, hydrogen peroxide offers a viable alternative for iron oxidation when overloaded facilities cannot be expanded, or where effluent pH limits can be met without pH adjustment (9).

5.3.4 Other Chemical Oxidants

In addition to ozone and hydrogen peroxide, chlorine and potassium permanganate are oxidants that could be applied to ferrous iron oxidation. Chlorine forms acid upon reaction with water, making it undesirable for AMD treatment. Both chlorine and permanganate are expensive chemicals; consequently, neither are economical for AMD treatment.

5.4 Biological Oxidation

Research indicates that bacteria capable of oxidizing ferrous iron exist naturally in most acid mine drainages. These bacteria, *Thiobacillus ferrooxidans*, obtain their carbon and nitrogen requirements from inorganic sources and their energy from the oxidation of ferrous iron. These bacteria have been implicated as a catalyst in the formation of acid mine drainage, and several methods have been proposed to utilize their ability to oxidize ferrous iron in mine drainage. Dispersed growth systems have proven unsuccessful; however, fixed growth systems, such as trickling filters and rotating biological contactors (RBC), are effective in supporting bacterial populations. Experiments by Lovell have shown that synthetic filter media best support the bacterial growth necessary for oxidation in a trickling filter, while rock material is subject to degradation by the acid water, causing clogging and ponding (10, 11).

Olem and Unz have completed research on ferrous iron oxidation of mine waters directly from the mine using rotating biological contactors (12, 13, 14). The RBC is an aerobic treatment device, consisting of a series of four plastic discs mounted on a horizontal shaft (see Figure 5-5). This assembly is placed in a trough through which the wastewater flows, submerging slightly less than half the surface area of the discs. The discs rotate slowly on the shaft, causing the biological growth on the discs to alternate contact between air and water. This is the first application of rotating discs to AMD treatment.

Olem and Unz obtained data from RBC pilot plants utilizing actual acid mine drainage. They compared performance at two peripheral disc velocities, 0.32 and 0.17 m/s (63 and 34 ft/min), and at five hydraulic loadings ranging between 110 and 440 $l/d/m^2$ (2.7 and 10.8 $gal/d/ft^2$). They found a linear relationship between ferrous iron removal and stage retention time (time per individual compartment). This relation held for any set of operating conditions. Also, a faster disc velocity produced better iron oxidation at any constant hydraulic loading. At a given disc velocity, an increase in hydraulic loading from 110 to 440 $l/d/m^2$ (2.7 to 10.8 $gal/d/ft^2$) resulted in an increase in effluent ferrous iron concentration, even though the ferrous iron oxidation rate also increased. The RBC can be expected to produce an effluent containing less than 10 mg/l ferrous iron at loading rates up to 88 kg Fe^{++} applied/d/1,000 m^2 of disc surface (18 lb Fe^{++}/d/1,000 ft).

More recent tests have been conducted on acid mine drainages that flow overland for more than a mile before reaching the treatment facility, thus exposing the water to stream ecology and ambient air temperatures. The data col-

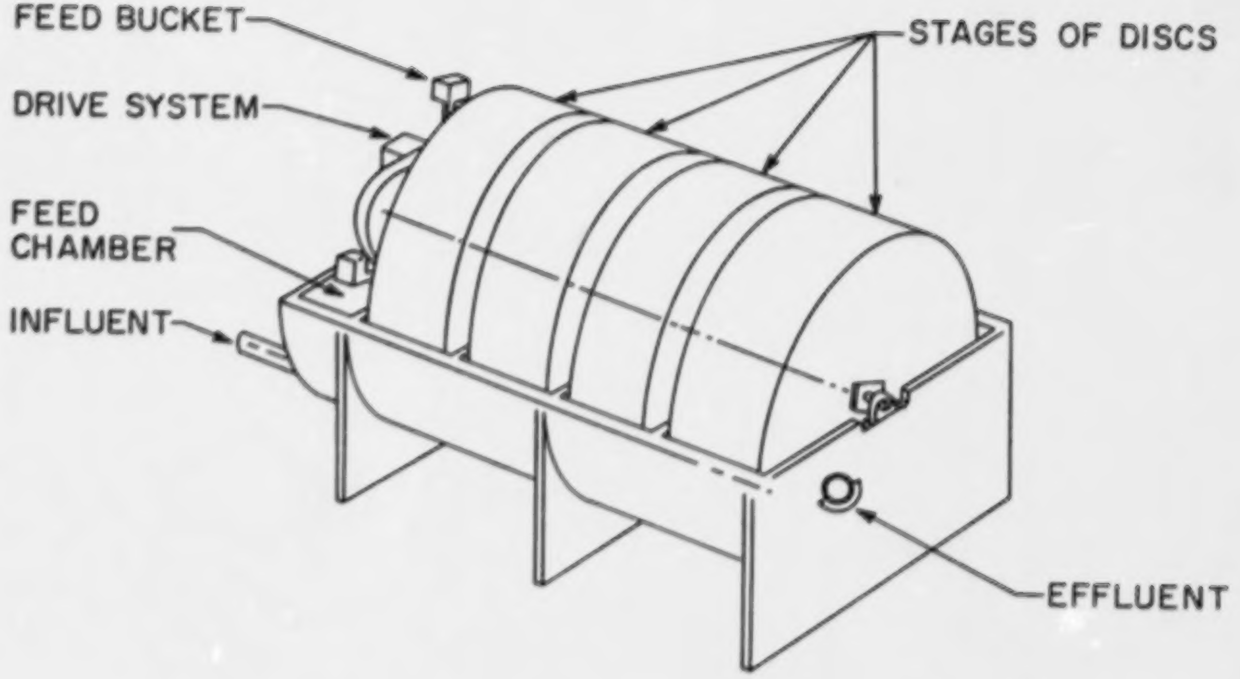
FEED BUCKET
DRIVE SYSTEM
FEED CHAMBER
INFLUENT
STAGES OF DISCS
EFFLUENT

Figure 5-5. Rotating biological contactor.

lected during winter operation showed that the effect of low temperatures on the biological oxidation process was not as great as expected. The removal efficiency at 0.4°C (32.7°F), the lowest temperature recorded, dropped only 10% below that found at 10°C (50°F), the initial mine water temperature. This effect can be negated by a lower hydraulic loading.

Although it is impossible to remove ferrous iron to 1.0 mg/l with the RBC, this degree of removal by a biological system is probably not necessary. Olem and Unz restate Lovell's belief that an effluent of 10 mg/l ferrous iron is adequate from the RBC. Since the effluent still must be neutralized, it is assumed that agitation during the neutralization step will provide enough aeration to oxidize any remaining ferrous iron.

Though the capital cost of the RBC system is greater than a conventional mechanical aeration system, the attraction of the RBC system is a significant reduction in energy costs through lower power requirements. The horsepower required to rotate the discs is substantially less than that required to power an aerator. The possibility arises that limestone can be used as the neutralizing agent, resulting in a savings in chemicals. Overall, the two most costly operational items (power and chemical) can be reduced. With the conventional aeration process, limestone neutralization is possible only in water containing less than 100 mg/l ferrous iron. Also, the RBC effluent pH need only be raised to 6.0, acceptable for discharge, while pH 8.5 is needed with the conventional system for efficient iron oxidation.

5.4.1 Design Procedure for a Four-Stage Configuration

Data obtained from the Olem and Unz study can be applied to the design of a four-stage single-shaft RBC system (12). Equation 19 is used to determine the required disc surface area.

$$A = \frac{FoQ\ 86.4}{L} \qquad (19)$$

where A = disc surface area (m^2)

Fo = initial ferrous iron concentration (mg/l)

Q = flow (l/s)

L = ferrous iron loading (kg/d/1,000 m^2) determined from Figure 5-6

$$A = \frac{FoQ}{120L}$$

where A = disc surface area (ft^2)

Fo = initial ferrous iron concentration (mg/l)

Q = flow (gal/d)

L = ferrous iron loading (lb/d/1,000 ft^2) determined from Figure 5-6

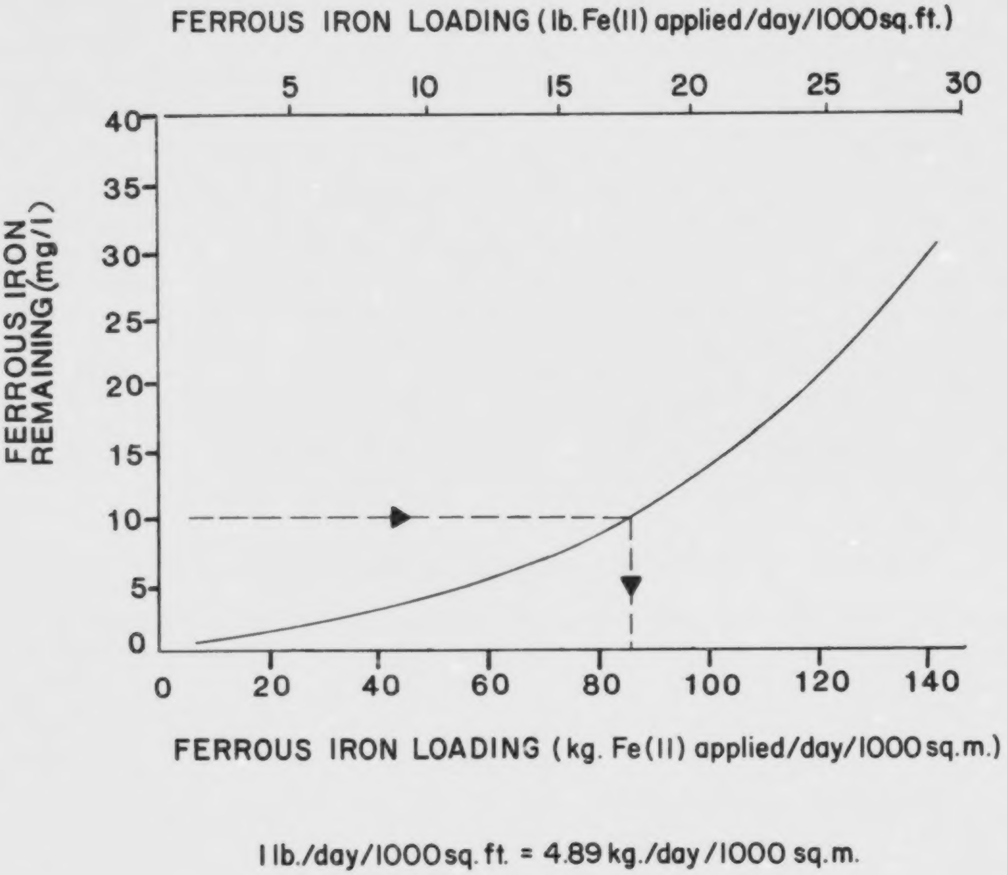

Figure 5-6. Design procedure for a four-stage rotating biological contactor configuration.

The designer chooses the desired effluent ferrous iron concentration from the RBC system listed on the vertical axis (Figure 5-6). A line is drawn horizontally, intersecting the curve, and dropped vertically from this point to the horizontal axis. Derived is the maximum ferrous iron loading (L) that will yield the desired effluent. This value, as well as the initial ferrous iron concentration (Fo) and the flow (Q), are applied to Equation 19, yielding the necessary disc area. RBC manufacturer data then must be utilized to determine the number of RBC discs needed to meet this surface area requirement.

5.5 Oxidation Rate Test Procedure

The oxidation of ferrous iron is a first-order reaction with respect to ferrous iron concentration, meaning that the reaction rate is dependent on the ferrous iron concentration at any point in time. This reaction rate can be expressed as

$$\frac{-d\ (Fe)}{dt} = k(Fe) \qquad (20)$$

where Fe is the ferrous iron concentration at any given time and k is the oxidation rate constant. The integrated form of the equation is

$$Fe = Fe_o e^{-kt} \qquad (21)$$

where Fe_o is the initial ferrous iron concentration and Fe is the ferrous iron concentration at time t.

This value of the reaction rate constant k, can be determined from a graph of log Fe vs. time. Ferrous iron oxidation data will yield a straight line when plotted on semilog paper. The slope of the line multiplied by 2.303 equals the rate constant, k.

$$k = \text{slope} \times 2.303 \qquad (22)$$

Once the rate constant for a particular mine drainage at a given operating pH is determined, the detention time required for oxidation of the same water from any initial concentration to any final concentration can be determined by Equation 23.

$$t = 1/k \times 2.303 \log\frac{(Fe_i)}{(Fe_f)} \qquad (23)$$

where Fe_i and Fe_f are the initial and final ferrous iron concentrations, respectively.

The oxidation rate test should be performed on fresh samples of mine drainage to minimize any natural oxidation of the ferrous iron. A large sample, preferably 4 l (1 gal) or so, should be used. The sample should be stirred continuously.

Initially, a small sample is taken and properly preserved for analysis to determine the initial ferrous iron concentration. Lime or caustic soda is added quickly to raise the pH to the desired operating level. At the same time, air is supplied through a diffuser such as an aeration stone or disc. Samples are then removed periodically throughout the test for ferrous iron analysis. All samples must be preserved properly with hydrochloric acid to prevent the oxidation from continuing. The pH should be measured frequently and maintained at the selected level.

A 1-hour test should be sufficient to generate data for determination of the oxidation rate constant. The test should be repeated once or twice to assure repetitive results.

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CHAPTER 6

SEDIMENTATION

6.1 General Characteristics of Mine Drainage Sludge

Mine drainage sludges are generally composed of hydrated ferrous or ferric oxides, gypsum, hydrated aluminum oxide, unused lime, varying amounts of sulfates, calcium carbonates, bicarbonates, and trace amounts of silica, phosphate, manganese, copper, titanium, and zinc (1). Hydroxides are formed by the reaction of metal salts with hydroxyl ions from natural or added alkalinity (lime, caustic).

The sludge characteristics vary with mine drainage quality and neutralization method. Important sludge characteristics include settleability, density, dewaterability, particle surface properties, and viscosity (2). Sludge settleability combines the aspects of sludge settling rate and final sludge volume. Sludge density is usually reported as percent solids by weight. It is extremely desirable to produce a dense sludge to reduce sludge volumes and handling and disposal costs. Sludge dewaterability is defined as the ability of a sludge to be concentrated into a more manageable and less voluminous form by centrifuging, filtering, or lagooning. The electrostatic charge, or particle surface property, is important in determining how well individual particles will flocculate into larger particles and settle out of suspension. The viscosity of sludge measures the flowability, sometimes an important consideration when pumping sludges.

Mine drainage neutralized with lime (hydrated or quicklime) produces sludges that settle slowly. These hydrate-produced sludges are light, gelatinous, and very voluminous.

The ferric hydroxide sludges associated with mine drainages have high iron concentrations and are generally a fluffy mass with very low solids, usually less than 1%. Sludge production increases as the iron and aluminum concentrations increase, averaging 5%-10%, and can be as high as 30% of the total treated volume. Sludges formed in the pH range of 6.4 to 7.2 have the best settling properties (2), but the iron oxidation rate is poor. The ideal pH for iron oxidation is around pH 8.5.

Sludges generated from mine drainage neutralized with limestone have a higher density than those generated from other lime products. The volume of limestone sludge can be as little as 20% that of sludge neutralized using lime, and the solids content can be up to 15 times greater. This results in a smaller volume required for sludge disposal. Sludges produced from highly acidic waters neutralized with caustic soda (NaOH) have very low densities and resist compaction despite long storage detention times. Even the low acidic waters, when neutralized with caustic, will produce large sludge volumes. Sludges formed from soda ash (sodium carbonate, Na_2CO_3) neutralization compact to densities somewhere between those generated by lime and limestone.

Sammarful (3) reported that mine drainage neutralized with carbonates yields a granular, dense sludge, while hydroxide neutralizing agents produce a semigelatinous sludge. Figure 6-1 illustrates the relationship of settling time to sludge volume for various neutralizing agents (4).

Polyelectrolyte addition may be used to improve sludge settling rates. Polyelectrolytes are water-soluble, high-molecular-weight, organic polymers that may be cationic (positively charged) or anionic (negatively charged). These charged polymers adhere to sludge particles and improve settling. Determinations about whether cationic and anionic polyeletrolytes should be used, and about what polymer dosages are proper, depend on sludge characteristics. Treatability tests provide a means of determining these variables.

Lovell (1) separated AMD sludge settling rates into three types or classifications of settling phenomena. Type 1 settling is associated with the neutralization of mine drainage containing high concentrations of iron and aluminum. The volume of the sludge (mostly ferric hydroxide) usually amounts to 5%-10% of the average daily treated volume. Type 1 sludge settles rapidly with a well-developed liquid-solid interface. This most common type of mine drainage settling phenomenon is illustrated by Figure 6-2 (5). In part I of the figure, the precipitated sludge is a homogeneous mixture throughout the sample. In II, stratification begins and the particles in Zone D (bottom) begin settling onto already settled particles. Water becomes trapped inside the layers, forming a gelatinous mass. The adjacent upper layer (Zone C) is a transition zone characterized by a suspended solids concentration lower than Zone D but greater than Zone B. The supernatant, Zone A, develops as the liquid-solids separation is completed. Settling continues, as illustrated in III, and Zones A and D increase in depth while B and C decrease. In IV, only two zone areas (A and D) remain, with a large majority of the solids present in Zone D. At this point, Zone D begins to compress. Compaction forces exerted by individual particle density cause the bridging of floc to break down as the structure is overcome by its own weight. The weakly trapped water is displaced by the weight of the sludge as it achieves equilibrium with the compressive strength of the floc. This volume occupied by the sludge usually represents the final volume.

Type 2 settling pertains to sludge generated from mine drainage containing low concentrations of iron and aluminum with a pH of 6.5-7.5 and little or no acidity. Lime addition is not always required. Since this class of mine drainage lacks nuclei for flocculation, the sludge (natural suspended solids) experiences poor settling, and most tends to remain in solution. The particles that do settle are light and fluffy in character and produce sludges with only 0.5% solids (6). Satisfactory liquid-solids separation may take several days, or perhaps not occur at all. If treatment does not require neutralization, the volume of sludge is totally dependent upon the iron content, suspended solids, and other possible precipitable elements present

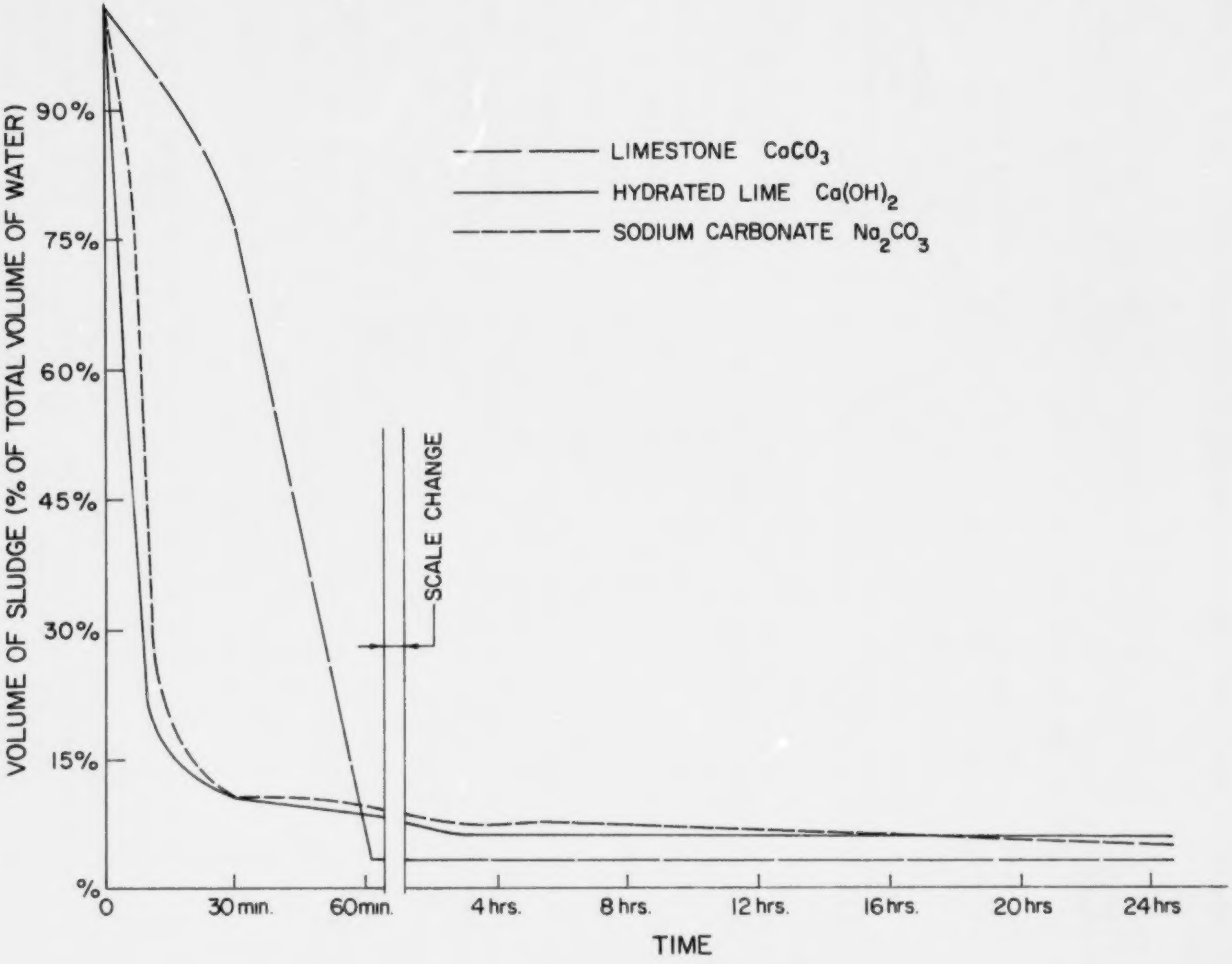

Figure 6-1. Treatability test settling curves.

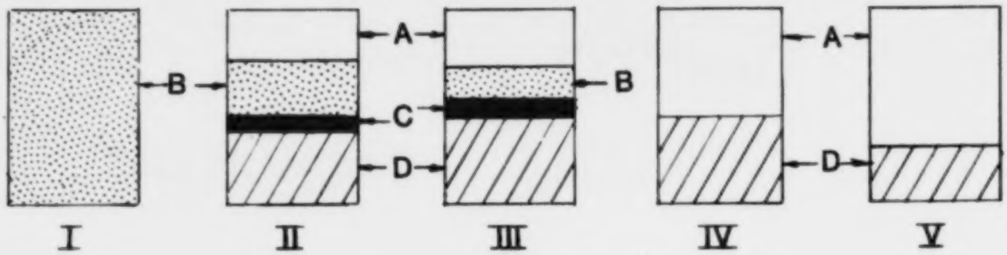

Figure 6-2. Type 1 settling.

in the water (7). Polymer or flocculant addition is highly recommended to enhance settling.

Type 3 settling behavior is independent of solids content and pollutional loadings and is generally associated with limestone or sodium carbonate (soda ash) neutralization. A two-phase separation system develops, with 90% of the solids settling rapidly. The solids settle as a cloud of individual particles without a distinct liquid-solid interface. The sludge volume approaches the maximum, usually within 10 minutes, yielding a turbid, cloudy supernatant. The supernatant turbidity can be controlled with the use of a mechanical upflow clarifier in which the influent passes through a sludge blanket. The sludge blanket filters the solids from the drainage, enhancing removal.

6.1.1 Settling Performance

Settling performance is related to the hydraulic surface loading, which is calculated by dividing the design flow (l/d) by the surface area of the pond or clarifier (m^2), with the resulting hydraulic surface loading in units of $l/d/m^2$ ($gal/d/ft^2$). Common values range from 175 to 350 $l/d/m^2$ (500 to 1,000 $gal/d/ft^2$).

Hazen showed that separator performance is related to hydraulic surface loading and the following variables (8):

1. flow turbulence in the basin;

2. velocity distribution throughout the pond;

3. particle interaction;

4. particle resuspension.

Starting at the water surface, a particle must settle the depth (D) of the separator at a velocity (Vs) such that the settling time is less than or equal to the time the liquid is in the basin (9). Expressed empirically,

$$\frac{\text{Depth (D)}}{\text{Settling Velocity (Vs)}} = \text{Detention Time} \tag{24}$$

$$\frac{\text{D (m)}}{\text{Vs (m/min)}} = \frac{\text{Volume } (\text{L} \times \text{W} \times \text{D, m}^3)}{\text{Flow } (\text{m}^3\text{/min})}$$

Simplifying,

$$\text{Vs (m/min)} = \frac{\text{Flow } (\text{m}^3\text{/min})}{\text{Surface Area } (\text{W} \times \text{L, m}^2)}$$

Therefore, all particles having a settling velocity (Vs) greater than or equal to the hydraulic surface loading (Flow/Area) are removed.

Actual basin performance is affected by many factors. Inlet and outlet devices and wind induce turbulent, nonquiescent flow currents that lead to poor settling and possible short-circuiting. Flocculation of particles into large agglomerations results in nonuniform settling velocities. Also, influents with high suspended solids concentrations tend to settle as a mass rather than as discrete particles.

6.2 Settling Unit Design

Settling units used in mine drainage treatment vary in size, configuration, and method of solids removal. Earthen settling ponds are the most popular because of their low capital and operational costs. The use of mechanical clarifiers or thickeners, however, should be considered because they enable the operator to exercise more control over the treatment system and improve sludge densities (thus yielding lower volumes). Mechanical separators will be discussed in detail later in this chapter.

A settling unit can be divided into four effective zones (see Figures 6-3 and 6-4) (10):

1. the inlet zone, in which the influent enters the unit and is uniformly distributed over the cross-sectional area;
2. the settling zone, where the majority of liquid-solid separation occurs;
3. the sludge zone, used for temporary or permanent storage and compaction of settled solids;
4. the outlet zone, where the supernatant is removed from the unit.

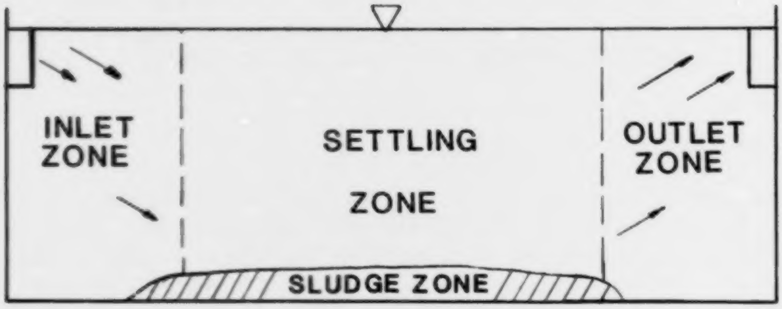
INLET ZONE
SETTLING ZONE
OUTLET ZONE
SLUDGE ZONE

Figure 6-3. Zones in a horizontal continuous flow sedimentation basin.

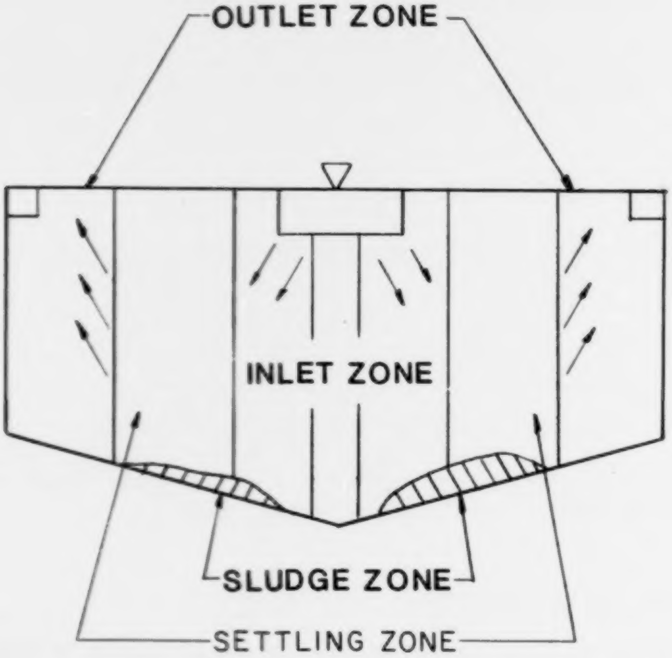
OUTLET ZONE
INLET ZONE
SLUDGE ZONE
SETTLING ZONE

Figure 6-4. Zones in a circular center feed, horizontal continuous flow sedimentation basin.

Little if any settling occurs in the inlet and outlet zones, reducing the effective settling area to Zone 2. An optimum settling unit design is one that minimizes the effects of the inlet and outlet zones.

6.2.1 Treatability Test

Before sizing and designing any settling pond or clarifier, a treatability study should be performed to determine the behavior and characteristics of the sludge along with expected supernatant quality. A simple testing procedure yields much information on the properties of the sludge that are important to design. These properties include sludge settling velocity, optimum pH, best neutralizing agent, dosage rate, and sludge density and volume. A recommended procedure for a treatability test can be found at the end of this chapter.

Results obtained from laboratory tests, while useful, cannot be applied directly to basin design. Sludge settling velocity, for example, which is extremely important in clarifier design, can be determined, but the value obtained is usually reduced to account for variables in the system.

Treatability studies also give an indication of sludge volume produced per unit volume of influent treated. This value is very important in determining tank volume.

6.2.2 Earthen Ponds

The following criteria should be considered when evaluating a site and constructing a settling pond.

6.2.2.1 Site Location

Earthen basins should not be located in swamps, marshes, or floodplains, on steep slopes, or over abandoned wells or mine workings which might have fissured rock that would permit seepage. Test borings are necessary when information on soil conditions is unavailable (11).

6.2.2.2 Soil Conditions

Ponds should contain a 0.6- to 0.9-m (2- to 3-ft) layer of impervious material to prevent seepage. Clays and silty clays are excellent for this purpose. Sandy clays are usually satisfactory. Coarse textured sands, sandy gravel mixtures, and gob materials are highly pervious, and therefore usually unsuitable. Limestone areas are especially hazardous as pond sites because of the possible presence of crevices or sinkholes in the limestone (12).

Soil types can be determined by laboratory testing of samples taken from the field or from local Soil Conservation Districts.

6.2.2.3 Foundation Conditions

Soil conditions must provide stable support for pond enbankment foundations, as well as the necessary resistance to passage of water. Good materials, providing stability and being impervious, are mixtures of coarse and fine textured soils, such as gravel-sand, sand-clay, and sand-silt.

Tight contact between embankment and foundation should be insured to control seepage properly along the plane of contact (13). One method for achieving tight contact is to use layers of clays or silty clays in the construction of the embankments. Synthetic liners or bentonite addition can also be used to seal embankments.

The basin bottom should not be founded on bedrock or on stony, rocky soils. The most suitable bottom foundation consists of a thick layer of relatively impervious consolidated material.

6.2.2.4 Embankment Construction

The embankments forming the sidewalls of the basin should be constructed of impervious materials similar to those used in the basin's liner. They should be placed in layers and properly compacted with a sheepsfoot roller. The side slopes depend on the properties of the fill and on the strength and stability of the foundation material. Average slopes for inside embankments are 2.5:1 or 3:1. Outside slopes should be gentle enough to allow a mower to cut grass safely (3:1 or greater), and should be protected from erosion by low-growing grass. The recommended minimum crown widths for earth embankments of various heights are shown in Table 6-1 (12). If the crown is to be used as a roadway, it should be at least 4.3 m (14 ft) wide at any height.

6.2.2.5 Design of Ponds

The settling pond should have an impermeable layer of clay or a plastic membrane liner in areas where fill material is unsuitable (permeabilities greater than 10^{-6} cm/s). A minimum freeboard of 0.61 m (2 ft) should be maintained in the pond to prevent overflow.

The inside slope of the embankments should be protected against erosion caused by wave action. Placing a 0.61-m (2-ft) wide collar of riprap (aggregate greater than 5 cm (2 in) in diameter) at the expected water level on the embankment is an effective method of erosion protection. Ponds having wide fluctuations of water levels require wider collars, 0.91-1.22 m (3-4 ft). Synthetic collars (liners) may also be used.

Inlet and outlet design is a critical factor in providing quiescent conditions for good settling pond performance. Inflow should be uniformly distributed over as much of the pond width as possible. This can be accomplished with the use of multiple inlets or a continuous width, multiple V-notch box weir. This decreases the inlet flow velocity, which reduces the probability of washout or resuspension of solids. Bad inlet design can lead

TABLE 6-1

RECOMMENDED MINIMUM CROWN WIDTHS

Height of Embankment		Minimum Crown Width	
m	(ft)	m	(ft)
1.52	(5)	0.912	(3)
1.52 - 3.04	(5 - 10)	2.43	(8)
3.04 - 4.57	(10 - 15)	3.04	(10)
4.57 - 6.1	(15 - 20[a])	3.65	(12)
6.1 - 7.62	(20 - 25[a])	4.26	(14)

---

[a] A 3.04-m (10-ft) wide horizontal bench is required for every 6.1 m (20 vertical ft) of embankment with 1%-3% backslope.

to short-circuiting, which reduces the detention time and removal efficiency of the settling pond. It also produces "dead" areas of noncirculating water in the settling ponds, creating channels of flow within the pond, as shown in Figures 6-5 and 6-6.

Uniform distribution of influent across the width of the settling pond inhibits isolated mounding of settled particles within the basin, thus maximizing sludge storage volumes.

Similarly, outlet devices should also be multiple or continuous to maintain low exit velocities. High exit velocities--that is, those greater than 0.304 m/s (1.0 ft/s) at the effluent--create turbulence that can resuspend settled solids, causing deterioration of effluent quality.

Baffles, selectively placed in a settling pond, prevent short-circuiting. Two types used are surface baffles and submerged baffles.

Surface baffling prevents short-circuiting by diverting the flow from a straight line across the pond to a less direct pattern, thus allowing for more uniform influent distribution. Surface baffles generally float on the pond surface and are anchored to the embankments with cables. They project a few centimeters above the water level, and downward 0.6-0.9 m (2-3 ft), or as much as one-half the pond depth. They can be constructed of flexible rubber, PVC, nylon, or wooden planks.

Submerged baffles are used for sludge containment. The baffles extend upward

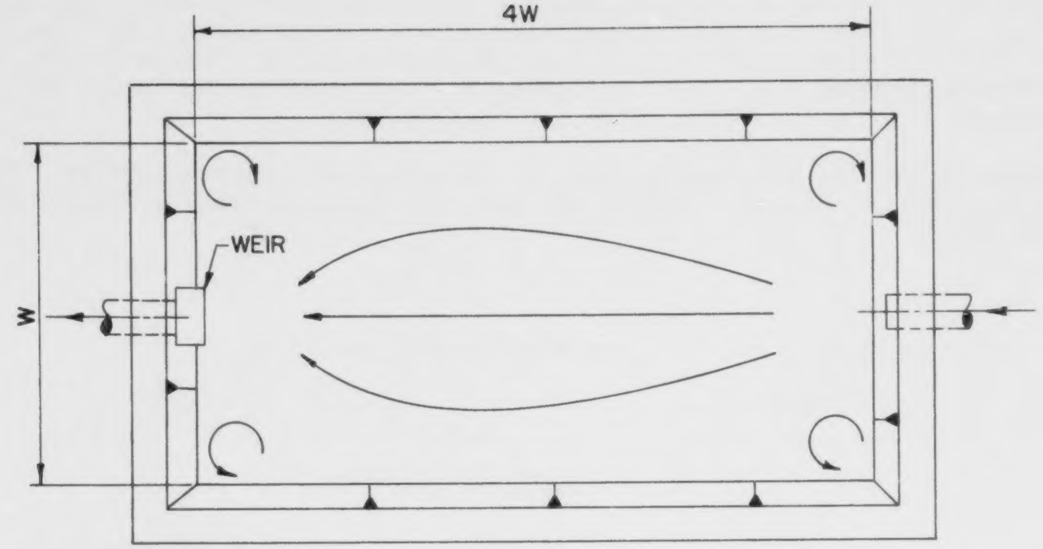

Figure 6-5. Nondistributed short-circuiting influent.

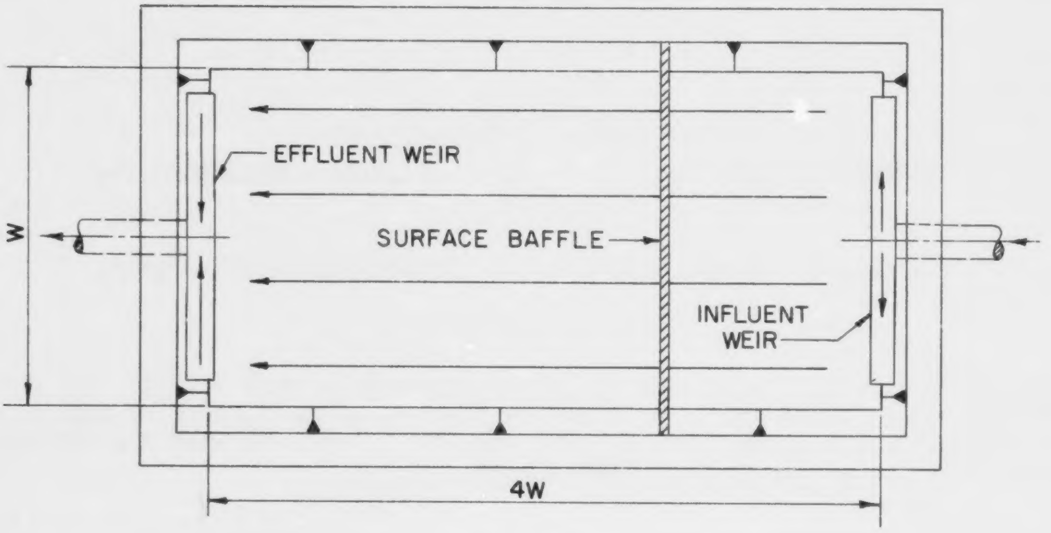
EFFLUENT WEIR
W
SURFACE BAFFLE
INFLUENT
WEIR
4W

Figure 6-6. Distributed influent.

from the bottom of the pond, and reach within a few meters (feet) of the surface. They create two separate settling chambers in the basin. The primary, or influent, chamber collects most of the settled solids. The secondary, or effluent, chamber acts like a polishing pond and provides additional solids removal. The separate chambers ease sludge removal.

Submerged baffles are usually used in conjunction with surface baffles. The combination of the two causes an "S" type flow path, as shown in Figures 6-7 and 6-8.

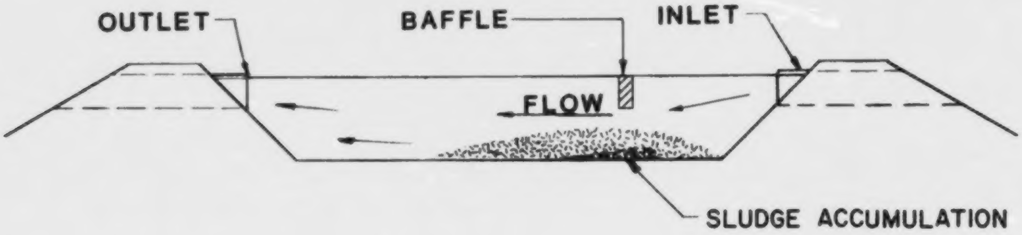

Figure 6-7. Surface-baffled pond.

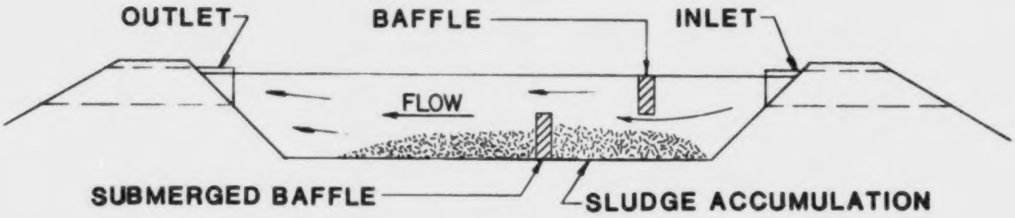

Figure 6-8. Combination surface- and submerged-baffled pond.

The effluent weir should not be placed on the leeward side of the pond because agitation caused by wave action from winds could cause resuspension of settled solids. The best effluent quality is achieved when the effluent weir is placed on the pond's windward side.

6.2.3 Types of Sedimentation Ponds

For the purpose of this manual, earthen ponds are separated into two general categories: settling ponds and impoundments.

Settling ponds are small and designed primarily for settling with periodic sludge removal. The sludge is disposed externally, or recycled to utilize the unreacted portions of the neutralizing agent and improve final density (solids content). These ponds require detailed engineering design and close operational control. There is usually more than one pond, and they are associated with more complicated treatment operations than impoundments.

Impoundments are large ponds designed for both settling and final sludge disposal. Sludge accumulates in the settling pond over the life of the mine, usually 10-20 years. Impoundments require large land areas, extreme depths, planned methods of handling surface runoff and drainages, and, usually, extensive construction.

To further define the characteristics of and design parameters for both basins, the following subsections are offered.

6.2.3.1 Small Earthen Settling Ponds (Approximately 2 Days Detention Time)

Small earthen settling ponds are designed to provide solids removal with limited sludge storage capacities (14). Ponds in this category generally have detention times less than 2 days (48 hours). These ponds require better initial design, construction, and quality operational control.

The optimum configuration for small earthen ponds is a rectangular shape with a length-to-width ratio of 4:1 (8). This long, narrow shape minimizes short-circuiting, and inlet and exit turbulence have minimal effect on the primary settling zone. Baffles, along with inlet and outlet distribution devices, are used extensively in these ponds. These ponds may be used either in series or in parallel in an effort to maximize efficiency and confine sludge accumulations.

A series operation allows the drainage to flow into the primary settling pond, which removes the majority of solids and is where sludge removal devices are usually located. The primary effluent then enters the second pond for final polishing.

A parallel pond system can operate in two ways. Neutralized water can flow to only one pond until the sludge accumulations begin carrying over into the effluent. At this time, the flow is diverted to the second pond until the sludge in the first is removed. The other method of operation uses both ponds simultaneously and requires dividing the flow with a splitter box. This operational mode usually employs sludge removal devices in each pond. The parallel mode provides a means of removing sludge without interrupting flow in the basins.

6.2.3.2 Impoundments (Greater Than 2 Days Detention Time)

Impoundments are designed to serve for both solids settling and final sludge disposal. They have detention times much greater than 2 days and are usually built as large as possible, without regard to detention time. Impoundments have an economic advantage over small ponds in labor and operating costs. Also, they provide the advantage of a large buffering capacity during treatment plant breakdowns, thus producing a more uniform effluent quality (15). Large impoundments, however, require extensive land area and are the least realistic conservation approach (16). More importantly, they are almost impossible to abandon.

Exceptionally large settling impoundments (detention time greater than or equal to 20 days) assume limnological properties of lakes, experiencing turnovers in both fall and spring. During summer, the water stratifies, forming a warm upper layer of freely circulating water (epilimnion); a middle layer (metalimnion) containing a rapid temperature drop with depth (thermocline); and a deep, cold, bottom layer (hypolimnion). As the sun decreases in intensity with the coming of fall, the water temperature and density become uniform throughout the impoundment. The slightest wind can now circulate the water, resuspending the sediments. A reverse effect occurs in the spring as the water restratifies, resulting in spring turnover (14). This entire phenomenon, however, is rare in larger impoundments.

This problem can be avoided by limiting the depth of the impoundment, enabling the sun to warm the entire body of water with a minimal temperature gradient. In many situations, however, this solution is impractical because impoundments must have the total volume for sludge disposal for the life of the mine, which can only be provided with excessive depths (greater than 6 m).

Large earthen impoundments are usually constructed by damming a valley or utilizing an abandoned strip pit. A special permit or state approval is sometimes required when impoundments obstruct watercourses or change watershed drainage patterns. Several states define embankments over certain heights as dams and require more stringent design and construction practices. When damming a valley, diversion ditches shuld be provided to route runoff around the impoundment. Runoff entering the impoundment increases the solids loading to the pond, decreasing sludge disposal capacity. Also, excessive amounts of runoff will cause resuspension of settled solids. Most importantly, eroded sediments can fill a basin quickly, greatly reducing the life of a facility.

The use of small impoundments (detention time 3-5 days) in parallel, each with sufficient volume to handle the total flow, can be effective and result in longer pond life. As sludge accumulation begins to reduce detention time in the first impoundment, the flow is diverted to the second. The sludge in the first impoundment is permitted to undergo compaction and drying by decanting the supernatant. The sludge dewatered atmospherically does not rehydrate and occupies a smaller volume compared to the original gelatinous precipitate (10). Ideal climatological drying will achieve a 30%-40% solids content; however, a sludge solids content of 10%-15% is satisfactory.

6.2.3.3 Volume Requirements

Many factors influence the sizing of sedimentation basins. These include detention time, sludge removal and disposal method, and mode of operation.

Detention time is the basic design parameter when sizing earthen settling ponds for mine drainage treatment, but it is not the only influencing factor. Inlet and outlet flow conditions must be considered, along with turbulence and eddy currents induced by wind. These disturbances reduce settling efficiency and resuspend partially or previously settled solids. Baffles reduce short-circuiting to a degree, but much less than the total basin volume is actually utilized. For these reasons, a minimum of 12 hours is required for sedimentation ponds utilizing periodic sludge disposal. A survey investigating performances of mine drainage treatment plants in southwestern Pennsylvania and northern West Virginia showed nearly all settling ponds had detention times in excess of 12 hours (17). It is common practice to build the ponds as large as possible, subject to land availability.

The basin volume can be calculated, using the detention time and design flow, with Equation 25.

$$V = Qt_d \tag{25}$$

where V = volume of settling pond, without sludge storage m^3 (gal)

Q = design flow, m^3/min (gal/min)

t_d = detention time (min)

Additional capacity in settling ponds must be provided for sludge accumulation. The sludge storage volume can either be approximated from methods explained previously or estimated conservatively as 5%-10% of the average daily volume treated.

Sludge storage volume requirements in a settling basin depend upon how often sludge is removed. Ponds constructed without sludge collection or removal devices should provide a minimum of 1-month sludge storage volume (11). Ponds equipped with sludge removal devices should allow sufficient volume for sludge storage between withdrawal operations. In some cases, basins are designed with enough volume to hold sludge for the life of the treatment plant, eliminating the need for any sludge removal. In all cases, the sludge storage volume must be included in the final volume of the basin.

Settling Volume + Sludge Volume = Pond Volume

6.2.4 Mechanical Clarifiers

Many mine drainage treatment plants employ clarifiers, generally circular, for liquid-solids separation. These devices are usually preferred when land area becomes limiting. The purpose of this subsection is to inform the designer of this alternative and to present criteria and cost data for economic and engineering evaluation.

When selecting a clarifier, the following features should be considered (18):

1. supernatant clarity;
2. an underflow of settled sludge having high density and suitable for secondary dewatering (a 3% solids content in the clarifier underflow is ideal (average 1%-2%));
3. provisions for sludge recycle to improve reagent utilization;
4. provisions for sludge removal to a second-stage thickening process (i.e., lagoons);
5. adequate sludge storage capacity (depth) for blanket formation, settling, and in situ concentrating;
6. provisions for flocculant addition to improve and control sludge settling rates.

Many existing mine drainage plants utilize one of the following three types of liquid-solids separators: (1) conventional clarifiers, (2) upflow solids contact or flocculator clarifiers, and (3) thickeners.

6.2.4.1 Clarifiers

As best defined, a clarifier is a gravitational liquid-solids separator having the primary objective of producing a high-quality supernatant (overflow) regardless of underflow solids content. Conversely, thickeners are used for the purpose of concentrating underflow, with secondary emphasis on supernatant quality. Consequently, most liquid-solid separators employed in mine drainage plants should be called clarifiers, but because of their large diameters requiring heavy-duty raking mechanisms, they are termed thickeners.

The first type of separator to be discussed is the conventional clarifier. More often employed in sewage treatment, this horizontal flow-type clarifier contains a rotating sludge removal mechanism with scraper blades, sludge hopper, drive motor and unit, and center feedwell. Clarifiers, as applied to mine drainage treatment, do not have skimmers. A conventional clarifier is shown in Figure 6-9.

The conventional clarifier operates much like a simple horizontal settling basin and utilizes radial flow distribution. Neutralized water enters the circular center feedwell that distributes the flow uniformly 0.6-0.9 m (2-3

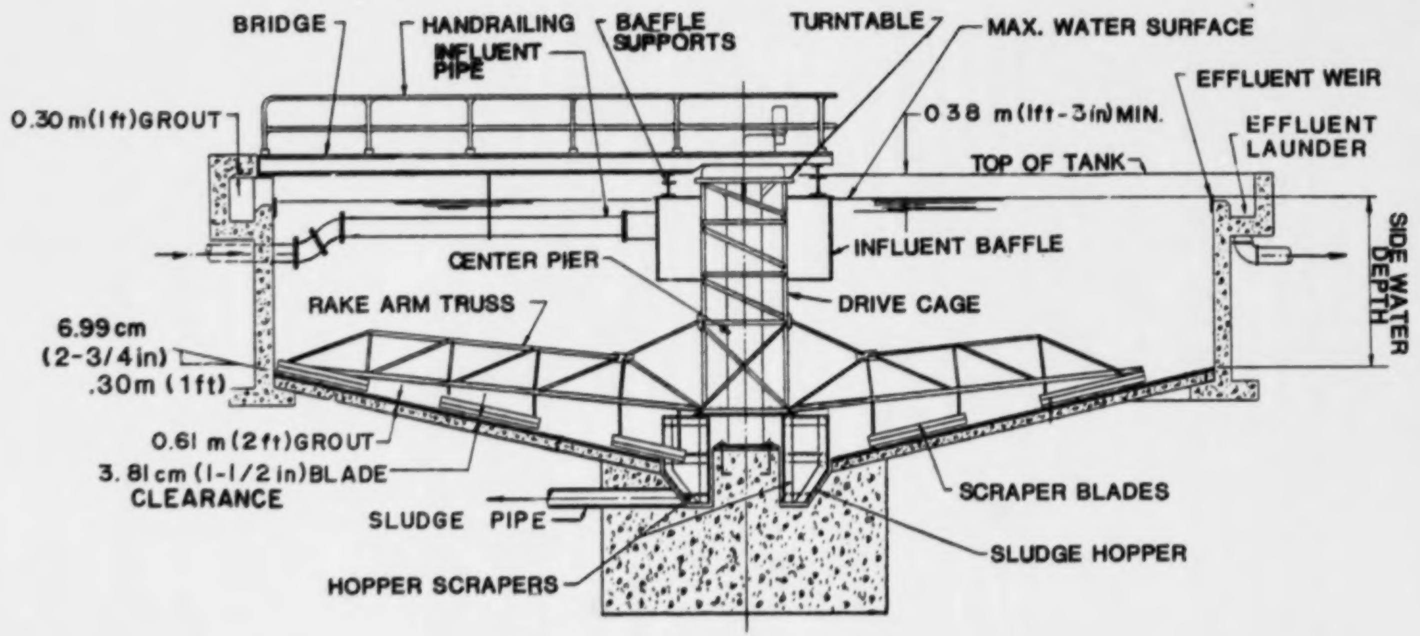
BRIDGE
HANDRAILING
INFLUENT PIPE
BAFFLE SUPPORTS
TURNTABLE
MAX. WATER SURFACE
EFFLUENT WEIR
0.30 m(1 ft)GROUT
0 38 m(1ft-3in)MIN.
EFFLUENT LAUNDER
TOP OF TANK
INFLUENT BAFFLE
SIDE WATER DEPTH
CENTER PIER
RAKE ARM TRUSS
DRIVE CAGE
6.99 cm (2-3/4 in)
.30m (1ft)
0.61 m(2 ft)GROUT
3.81 cm (1-1/2 in)BLADE CLEARANCE
SCRAPER BLADES
SLUDGE PIPE
SLUDGE HOPPER
HOPPER SCRAPERS

Figure 6-9. Conventional clarifier.

ft) below the surface. The water flows toward the periphery and the solids are separated by gravity. A continuous overflow weir is provided for supernatant outflow.

Sludge accumulations are moved to the collection hopper at the center by a rotating rake mechanism equipped with collection blades. The designer usually allows the volume required to cover the raking device as sludge storage. This depth is dependent upon clarifier diameter and mechanism selected. Sludge can be drawn off either continuously or intermittently. Higher underflow solids contents have been realized with intermittent drawoff.

The second type of clarifier, and perhaps the most popular unit, is the flocculator or upflow solids contact clarifier, as shown in Figure 6-10. This unit combines two functions into a single operation: flocculation (particle formation) and filtration. The flocculation occurs when a polymer is injected into the neutralized mine drainage in the feedwell. The flocculator paddles gently mix the stream, inducing floc formation. The skirted bottom on the feedwell extends below the sludge blanket level. This directs the newly generated floc to rise through the sludge blanket, where suspended solids undergo removal by adsorptive filtration. When this unit is designed and operated properly, a definite sludge blanket-supernatant interface can be seen within the clarifier.

More commonly, this unit is used without the addition of polymers for flocculation. This type of application is termed an upflow solids-contact clarifier. The floc formed during the neutralization reaction settles poorly, but well enough to create a filtration blanket. All principles of operation and flow patterns are the same as those of the flocculator-clarifier.

This mode of operation has produced high-quality effluents for plants utilizing either hydrated lime or quicklime as the neutralizing agent. The sizing of these clarifiers has principally been the responsibility of the vendor, and based upon one special parameter, rise rate. Rise rate is defined as the vertical velocity of the water in meters (feet) per minute or hour through the clarifier. For example, assume an existing plant has a daily average flow of 3,785 m^3 (1,000,000 gal) and a clarifier 30.48 m (100 ft) in diameter. The following calculations are performed using this standard formula:

$$V_{RR} = Q/A \qquad (26)$$

where V_{RR} = rise rate, m/min (ft/min)

Q = design flow rate, m^3/min (ft^3/min)

A = cross-sectional area of clarifier, m^2 (ft^2)

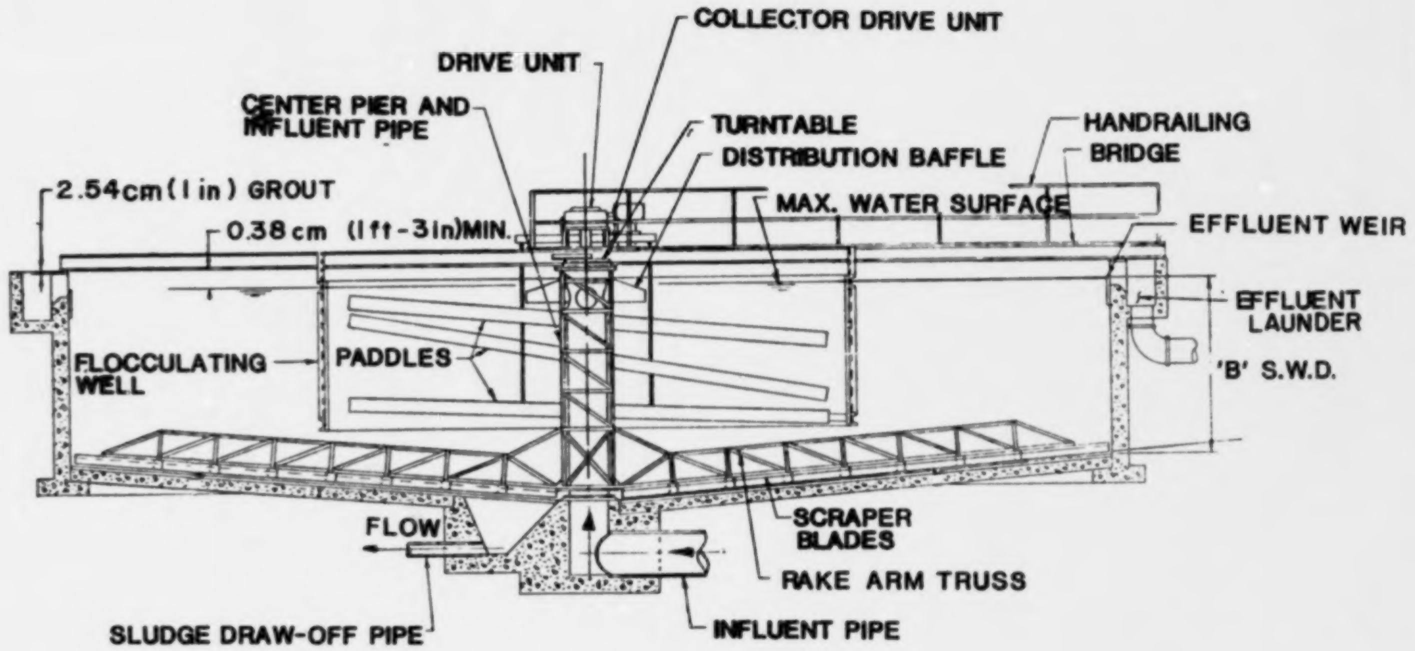
COLLECTOR DRIVE UNIT
DRIVE UNIT
CENTER PIER AND INFLUENT PIPE
TURNTABLE
DISTRIBUTION BAFFLE
HANDRAILING
BRIDGE
2.54cm(1 in) GROUT
0.38cm (1ft-3in)MIN.
MAX. WATER SURFACE
EFFLUENT WEIR
EFFLUENT LAUNDER
'B' S.W.D.
FLOCCULATING WELL
PADDLES
FLOW
SCRAPER BLADES
RAKE ARM TRUSS
SLUDGE DRAW-OFF PIPE
INFLUENT PIPE

Figure 6-10. Upflow flocculator clarifier.

	Metric	English
Flow:	$Q = \frac{3,785\ m^3}{d} \times \frac{d}{1,440\ min}$	$\frac{1,000,000\ gal}{d} \times \frac{ft^3}{7.45\ gal} \times \frac{d}{1,440\ min}$
	$Q = 2.63\ m^3/min$	$93.2\ ft^3/min$
Area:	$A = \frac{(30.48\ m)^2}{4} * \pi$	$\frac{(100)^2}{4} * \pi$
	$A = 729.28\ m^2$	$7,854\ ft^2$
Velocity:	$V_{RR} = Q/A$	
(rise rate)	$= \frac{2.63\ m^3/min}{729.28\ m^2}$	$\frac{93.2\ ft^3/min}{7,854\ ft^2}$
	$V_{RR} = 0.0036\ m/min$	$0.012\ ft/min$

Thus, if a rise rate is assumed, the formula can be rearranged to yield the appropriate surface area or equivalent diameter. The most commonly used rise rate (V_{RR}) in sizing a mine drainage clarifier is 0.015 m/min (0.05 ft/min). In actual practice, however, the values acquired from reliable performance plants are lower.

Illustrated in Table 6-2, items A-1 through A-12, are actual flows and sized clarifiers in operation. These particular plants produce effluents in compliance with EPA standards and with a high degree of reliance. As shown, the rise rates, with the exception of A-9, are below the vendor recommendation. This is not to say that a 0.015-m/min (0.05-ft/min) rise rate will undersize the clarifier, resulting in poor effluent quality. Industry has elected to oversize clarifiers in an effort to guarantee performance, and to allow additional capacity as a safety factor for higher-than-predicted drainage flows.

Based upon these data, the designer can size a clarifier with confidence, producing good performance when using a rise rate in the range of 0.003-0.009 m/min (0.01-0.03 ft/min).

The depth of the clarifier, be it solids-contact, flocculator, or conventional, is independent of surface area and depends upon the detention time, sludge blanket thickness, required storage volume, raking mechanism, and desired underflow percent solids content.

Lovell (18) lists various optimum design parameters, which ultimately determine depth in addition to size. As shown in Table 6-3, the suggested 1.83-m (6-ft) minimum sludge depth below the feedwell, plus a 72-hour minimum detention time for sludge storage, significantly influences the design depth.

Similar to values found in actual practice, 0.003-0.009 m (0.01-0.03 ft/min), Lovell (18) recommended rise rates of 0.136 $l/s/m^2$ (0.2 $gal/min/ft^2$) equivalent to 0.008 m/min (0.026 ft/min).

TABLE 6-2

RISE RATES FOR EXISTING MINE DRAINAGE CLARIFIERS

AMD Plant	Flow		Clarifier, dia		Rise Rate (V_{RR})	
	m^3/d	Mgal/d	m	ft	m/min	ft/min
A-1	3,785	1.0	18.9	62	0.0091	0.03
A-2	15,140	4.0	67.0	220	0.0030	0.01
A-3	26,495	7.0	67.0	220	0.0057	0.017
A-4	28,387	7.5	57.9	190	0.0073	0.024
A-5	16,275	4.3	57.9	190	0.0042	0.014
A-6	23,467	6.2	54.8	180	0.0067	0.022
A-7	25,738	6.8	54.8	180	0.0076	0.025
A-8	3,475	0.92	22.8	75	0.0057	0.019
A-9	11,355	3.0	22.8	75	0.0192	0.063
A-10	946	0.25	7.6	25	0.0143	0.047
A-11	3,785	1.0	30.5	100	0.0033	0.011
A-12	18,925	5.0	45.7	150	0.0079	0.026
Vendor recommendation	--	--	--	--	0.0152	0.05

At this point, the designer can estimate clarifier size and realize the land area requirements. Knowing the diameter, a rule of thumb for cost is $5,000/m ($1,500/ft) of diameter excluding the shell. (See Chapter 13, Cost Estimating, for details.) Commonly, the shell is constructed totally of concrete, but steel has been used in combination with it.

6.2.4.2 Thickeners

Thickeners are those separator units that usually receive clarifier underflow or an influent with a high percent solids. For example, the separators associated with coal preparation plants are true thickeners. These units are equipped with heavy-duty raking mechanisms and receive influent (washwaters) high in suspended solids. Consequently, most separators used by the coal industry, such as those designed for mine drainage treatment, have been termed "thickeners," which may not always be correct.

Nevertheless, thickeners have been used extensively in mine drainage treatment to perform the following double function: (1) to produce an overflow within federal and state standards, and (2) to store sludge to produce denser underflows.

Thickeners are sized like clarifiers, but design parameters depend upon the intended purpose of the unit. If a thickener is the only separator unit before ultimate disposal, the designer may consider a longer-than-normal (12-hour) detention time in an attempt to produce a dense underflow. These

TABLE 6-3

SUGGESTED OPTIMUM DESIGN PARAMETERS FOR CLARIFIER OPERATION

1.	Rising water velocity, $l/s/m^2$ ($gal/min/ft^2$)	0.136 (0.2) maximum
2.	Solids settling rate, m/hr (ft/hr)	0.61 (2.0) minimum
3.	Unit area, $m^2/kkg/d$ ($ft^2/ton/d$)	30 (300) minimum (may exceed 1,024 (10,000))
4.	Percent solids in underflow	5.0 minimum
5.	Sludge recycle provision - percent circulating load	5-30
6.	Percent solids to feedwell (combined raw feed plus recycle sludge)	1 minimum
7.	Flocculant dosage, mg/l influent	1 - 2
8.	Depth of feedwell below sludge level, m (ft)	0.61 (2) minimum
9.	Sludge depth below feedwell, m (ft)	1.83 (6) minimum
10.	Settled sludge retention time, hr	72 minimum

units are of large diameter, usually greater than 30.48 m (100 ft), depending upon design flow.

Smaller diameter thickeners are used as secondary separators, receiving clarifier underflow or settled sludge from earthen basins. Operating under this condition, the primary purpose of the unit is to thicken or dewater sludge. Usually, special flocculants or polymers are added to enhance the process.

Regardless of the design purpose, thickeners should be sized using a rise rate between 0.003 and 0.009 m/min (0.01 and 0.03 ft/min). This rate is more conservative than the vendor recommendation of 0.02 m/min (0.066 ft/min), but can vary with the technique used for sludge witdrawal (periodic or continuous).

Also, thickeners can be sized on the basis of solids loading. Surface area, expressed as $m^2/kkg/d$ ($ft^2/ton/d$) can range from 4.09 to 49.16 $m^2/kkg/d$ (40 to 480 $ft^2/ton/d$), depending upon the total process operation. The lower values are applicable when either polymer or sludge recycle is employed, while the higher values apply for a once-through system utilizing gravity separation only. An average design value of 30 $m^2/kkg/d$ (300 $ft^2/ton/d$) is often used. Only a treatability test simulating actual operating conditions, however, can produce a valid loading rate.

6.2.4.3 Supplemental Operational and Design Considerations

The operational method of a clarifier or thickener is usually the result of on-site adjustments that produce the best effluent quality. Every treatment plant has its own peculiar behavior that deviates from laboratory results.

For example, the raking mechanism of a separator can be operated continuously or intermittently. Best results have been obtained when the collection rake operates periodically. This allows the settling sludge total quiescence for maximum compaction. After a period of time (usually days), the collection rake is activated and sludge withdrawn.

Plants producing large volumes of sludge may require the collection rake to operate continuously. In such cases, the designer could consider a slower-than-normal peripheral speed. Generally, a clarifier collection mechanism moves at 4.57-6.1 m/min (15-20 ft/min), while thickener rakes rotate slightly faster, 7.62-9.14 m/min (25-30 ft/min). Even slower peripheral speeds of 1.5-3.0 m/min (5-10 ft/min) can improve sludge densities; their use merits consideration.

Also, the support structure of the collecting rakes warrants forethought, especially for those plants anticipating a gypsum ($CaSO_4$) problem. Many types of clarifiers and thickener mechanisms are constructed with various features to allow easy repair or cleaning. Most of these underwater structures (trusses) are massive and designed as cantilevers. Thus, 0.454 kg (1 lb) applied to the rake arm 15.24 m (50 ft) from the center support exerts 6.92 kg-m (50 lb-ft) of torque. Therefore, small amounts of sludge accumulation will produce a high torque requirement that must be transferred through the raking mechanisms. Typical raking mechanisms plow through the sludge and slowly move it to the collection well. Where gypsum is not a problem, these mechanisms function satisfactorily.

If the chemistry of an acid mine drainage indicates a gypsum problem might occur (sulfate greater than or equal to 2,500 mg/l), these raking mechanisms with large underwater surface areas should be avoided. The alternative is a raking mechanism pulled by cables, as shown in Figure 6-11. The drive or torque truss is visible above water, and only the cables and raking blades are submerged. Obviously, this minimizes underwater surface area available for precipitating gypsum. Also, the cables are flexible supports from which thin layers of gypsum can be easily removed.

Mechanical separators function well with a minimum of maintenance. Concrete structures are preferred; however, steel shells can be used when provided with corrosion protection. The designer should allow for operator access to potential problem areas, such as the thickener rakes and sludge withdrawal line, without having to drain the unit.

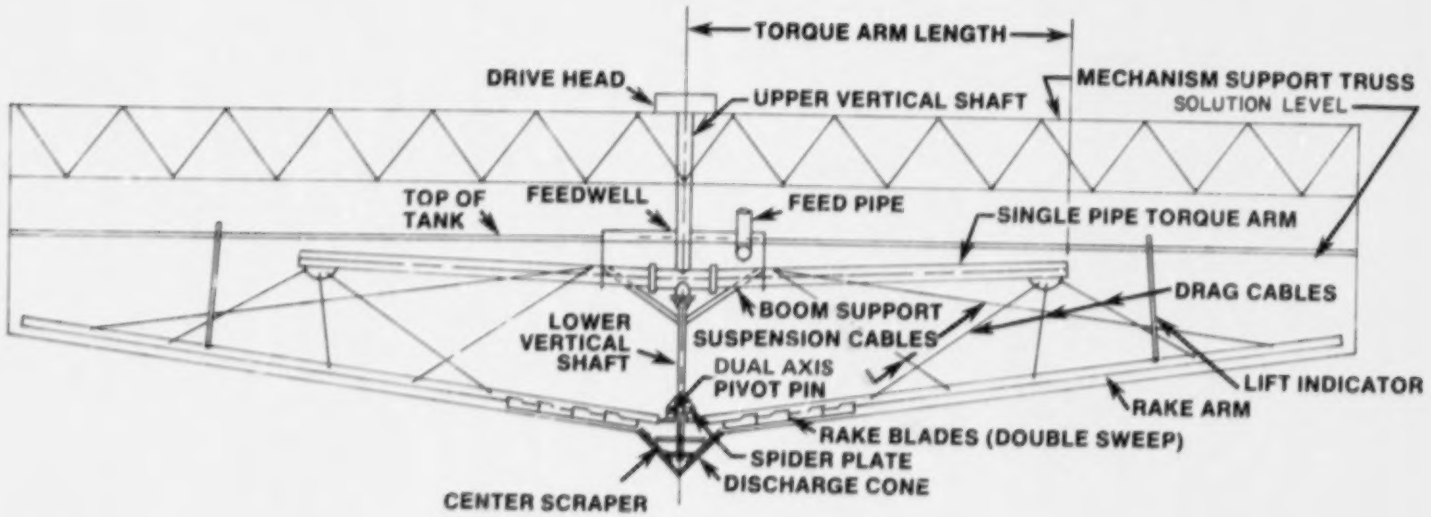
TORQUE ARM LENGTH
DRIVE HEAD
UPPER VERTICAL SHAFT
MECHANISM SUPPORT TRUSS
SOLUTION LEVEL
TOP OF TANK
FEEDWELL
FEED PIPE
SINGLE PIPE TORQUE ARM
DRAG CABLES
BOOM SUPPORT
SUSPENSION CABLES
LOWER VERTICAL SHAFT
DUAL AXIS PIVOT PIN
LIFT INDICATOR
RAKE ARM
RAKE BLADES (DOUBLE SWEEP)
SPIDER PLATE
DISCHARGE CONE
CENTER SCRAPER

Figure 6-11. Cable thickener.

6.2.4.4 Tilted-Plate Gravity Settlers

The tilted-plate gravity settler is an inclined-plate, shallow-depth settling device. It performs the same function as a conventional clarifier, but occupies approximately one-tenth the space. These settlers operate on the principle that solids removal is independent of depth and dependent upon the surface area of the settling unit. If the depth of the settling compartment is reduced to a few centimeters (inches) and a number of units are stacked, increasing available surface area, the handling capacity of the separator increases proportionately. These types of separator units use parallel plates tilted at an angle (55°-60°) so that the settled sludge is self-draining. The true effective surface area of the unit is calculated by using the cosine of the inclining angle and projecting the plates to a horizontal plane multiplied by 80%. Twenty percent of the area is lost to influent turbulence and collected sludge volume.

$$\text{Total surface area} = (N)\ (A \cos \alpha)\ (\text{width}) \qquad (27)$$

where N = number of plates

A = length of the plate (hypotenuse)

α = angle of inclination

This compacts the surface area of a clarifier into a much smaller size than those previously discussed.

The influent is introduced into the unit through a bottomless rectangular feed box located between sets of tilted plates. The influent flows into the plates from the side and then upward, exiting at the top of the tank through flow distribution orifices. The orifices are sized to take a specific pressure drop, insuring uniform flow distribution across the plates. The solids settle in each compartment and slide downward into the sludge hopper. Further concentration of the settled solids is accomplished by compression with a low-amplitude vibrator pack or with a specially designed version of a conventional thickener.

Figure 6-12 illustrates the flow pattern and compactness of the unit. Although to date few units are on-line in AMD treatment plants, tremendous potential exists for this type of application.

The criteria for sizing or computing the required surface area of the unit is based on 0.34 $\mathrm{l/s/m^2}$ (0.5 $\mathrm{gal/min/ft^2}$). This surface loading exceeds our recommended rise rate of 0.003-0.009 m/min (0.01-0.03 ft/min), but can be justified because most of these units are preceded by polymer flocculation.

The appropriate size unit is selected by equivalent clarifier diameter. For example, a process flow requiring a 13.7-m (45-ft) diameter clarifier would use a Model 2000/55, where 2,000 is the surface area ($\mathrm{ft^2}$) and 55° the angle of plate inclination.

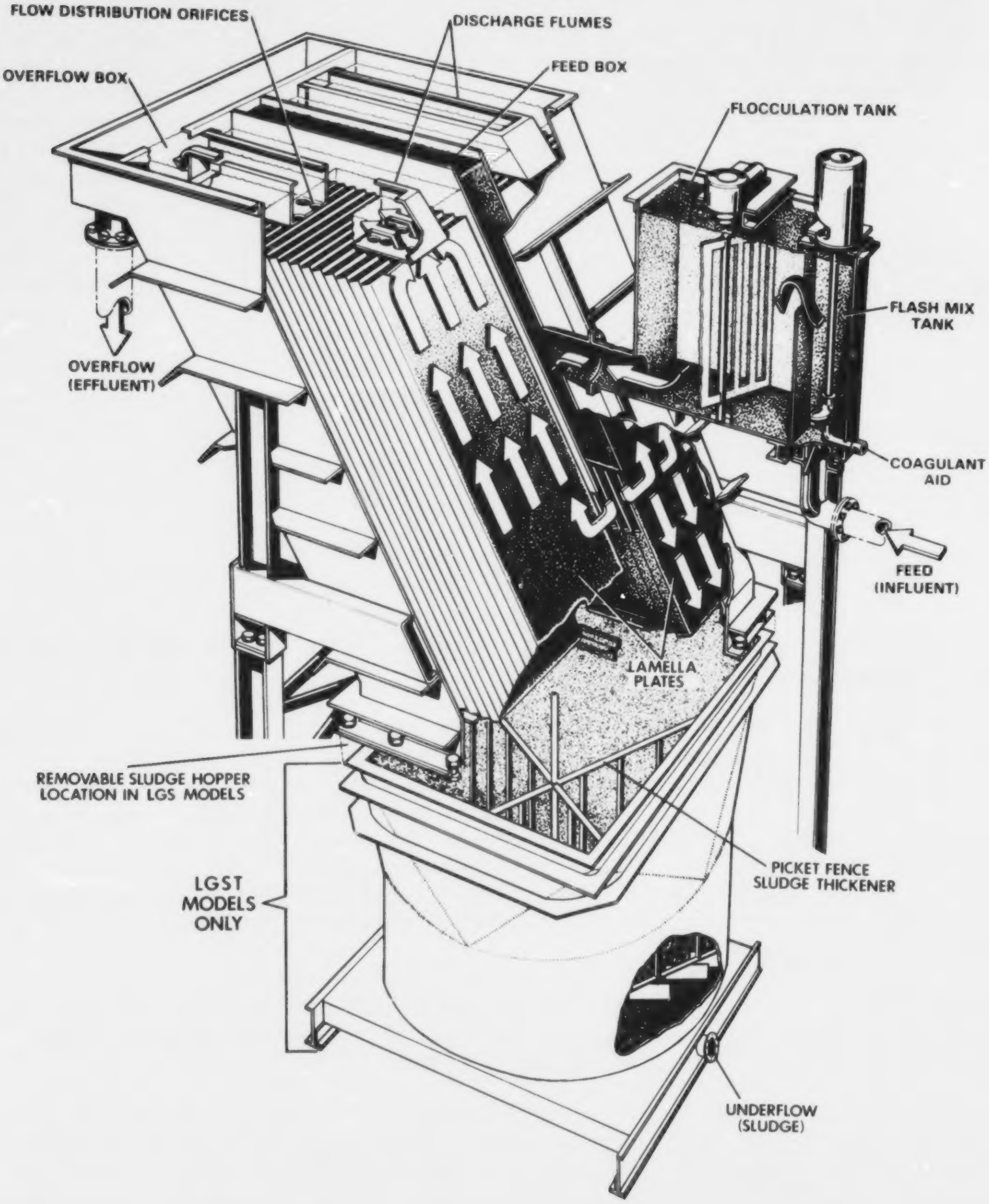
FLOW DISTRIBUTION ORIFICES
DISCHARGE FLUMES
OVERFLOW BOX
FEED BOX
FLOCCULATION TANK
FLASH MIX TANK
OVERFLOW (EFFLUENT)
COAGULANT AID
FEED (INFLUENT)
LAMELLA PLATES
REMOVABLE SLUDGE HOPPER LOCATION IN LGS MODELS
PICKET FENCE SLUDGE THICKENER
LGST MODELS ONLY
UNDERFLOW (SLUDGE)

Figure 6-12. Tilted-plate gravity settler.

The unit's fully assembled approximate cost can be estimated by multiplying the equivalent clarifier diameter by $5,000/m ($1,500/ft).

The major disadvantage of this type of separator is its susceptibility to clogging caused by gypsum. Where gypsum precipitation is suspected, this type of separator is not recommended.

6.3 Recommended Procedure for a Treatability Settling Test

1. Pour 1,000-ml portions of well-mixed, fresh (24 hours old or less) mine drainage into five 1,500-ml beakers. (If lab equipment is limited, only one sample at a time can be used.)

2. Place the beakers on a gang stirrer or comparable mixing device and stir for 5 minutes at 60-80 r/min.

3. Leaving one sample as a "blank" for comparison, add measured amounts of neutralizing agent to each sample, adjusting the pH levels between 6.0 and 9.0. Typically, if ferrous iron removal is the primary objective, a pH of 8.0 is strongly recommended. Agitate the mine drainage samples continuously during alkali additions, and thereafter for approximately 5 minutes at 80-90 r/min to insure thorough mixing.

4. Then, aerate each solution approximately 30 minutes either mechanically or with diffused air. Remove from each sample an aliquot for suspended solids analysis.

5. Pour each sample into a 1,000-ml graduated cylinder and allow to settle.

6. At 2-minute intervals, record the sludge interface height in the graduated cylinders for 30 minutes. Continue recording this height for the next 60 minutes at 10-minute intervals.

7. Plot the data recorded on a graph of liquid-solids interface height versus time for each sample tested. The plots should form curves similar to those shown previously in Figure 6-2.

8. Note the final sludge volume (mm). After acquiring the initial suspended solids concentration (mg/l) (step 4), the density of the final sludge can be calculated as follows:

$$ViSi = VsSs \qquad (28)$$

$$Ss = \frac{ViSi}{Vs}$$

where Vi = volume of sample (1,000 ml)

Si = initial suspended solids concentration before settling (mg/l)

V_s = final settled sludge volume (ml)

S_s = solids content of settled sludge

The calculation assumes 100% removal of all solids, which is not true. The supernatant will have some concentration of suspended solids; however, for a close approximation of the expected sludge density, the procedure is valid. Another method to use that ignores sludge density is the volume of sludge produced per volume of water treated. This ratio is the sludge volume ratio (14).

$$\text{Sludge Volume Ratio} = \frac{\text{Volume of Settled Sludge (ml)}}{\text{Volume of Sample (ml)}} \qquad (29)$$

The sludge volume ratio times the quantity of AMD treated per day yields a close approximation of expected sludge production per day. This method can be used if a laboratory analysis has been performed on the mine drainage to be treated. The sludge production can be estimated using the sum of the suspended solids and ferrous iron concentrations.

$$\text{Suspended solids concentration and dissolved iron concentration, mg/l}$$
$$\times \text{ flow, l/d} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = \text{sludge solids produced, kg/d}$$

Assume 15% of the neutralizing agent is waste and settles with the sludge.

$$\text{Sludge solids produced, kg/d} + 0.15 \text{ neutralizing agent added, kg/d}$$
$$= \text{total settled solids, kg/d}$$

Assume an average 1% solids content in the sludge or results from step 8.

$$\frac{\text{Total settled solids, kg/d}}{0.01} = \text{sludge produced, kg/d} \times \frac{1 \text{ l}}{1 \text{ kg}} \times \frac{10^{-3} \text{ m}}{\text{l}}$$
$$= \text{sludge produced, m}^3\text{/d}$$

9. The quality of the supernatant is determined by decanting portions of each sample tested and analyzing for effluent parameters (iron, pH, suspended solids, manganese). If the supernatant quality does not meet the standards of Environmental Protection Agency Effluent Guidelines and state standards, an extension of the detention time, use of an alternative neutralizing agent, or addition of polymer settling aids should be considered. If supernatant quality is obviously undesirable across the

pH range tested, then polymer addition becomes necessary. At this point, it is best recommended that a polymer salesman be contacted for the type and dosage of flocculant required.

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CHAPTER 7

SLUDGE DEWATERING AND DISPOSAL

7.1 Introduction

Neutralization of acid mine drainage creates a sludge that can be costly to handle, dewater, and ultimately dispose. Today, sludge handling and disposal presents the most recurrent and demanding problem to the designer or operator of a mine drainage treatment plant.

Environmental regulations coupled with all their ramifications for sludge disposal categorize sludge as a secondary or potential pollutant. In designing a mine drainage treatment plant, it is important to realize there are two effluents; i.e., treated water and sludge.

The volume of sludge from a mine drainage treatment plant varies with the composition of the untreated drainage and the neutralization method employed. However, a designer should expect the sludge volume to be from 5% to 10% of the daily flow through a typical treatment facility. Some facilities have recorded sludge volumes as high as 33% of the average daily flow (1). Therefore, the methods of sludge handling are extremely important to the designer.

The purpose of this chapter is to describe the chemical and physical properties of mine drainage sludge, the practical methods of dewatering, and to discuss the most commonly employed sludge disposal methods.

7.2 Mine Drainage Sludge

This section provides a description of mine drainage sludge, including its chemical characteristics and physical properties, such as its settleability, density, dewaterability, particle characteristics, and viscosity.

7.2.1 Chemical Characteristics

The chemical composition of AMD sludge, or yellowboy, varies with the raw drainage and method of treatment. Lovell reports that the sludge is generally composed of hydrated ferrous or ferric oxides, gypsum (calcium sulfate), hydrated aluminum oxide, varying amounts of sulfates, calcium, carbonates, bicarbonates, and trace quantities of silica, phosphate, manganese, titanium, copper, and zinc (2).

Illustrated in Table 7-1 are the primary components and compounds of severa

TABLE 7-1

CHEMICAL ANALYSES OF SLUDGES

Weight, % (dry basis 105°C-24 hr)

Alkali Used	Hydrated Lime - Air Oxidation			Hydrated Lime Bio-oxidation	Hydrated Dolomite	Calcined Dolomite		
Mine Water	**Bennett's Branch**	**Proctor 1**	**Proctor 2**	**Proctor 2**	**Proctor 2**	**Tyler**	**Proctor 1**	**Proctor 2**
Component								
Al	3.8	4.7	3.1	8.0	2.8	5.5	4.5	4.8
Fe	19.5	17.7	23.1	24.3	13.0	7.4	13.5	23.2
Ca	6.9	5.8	5.2	4.8	17.2	10.7	6.7	5.2
Mg	6.6	4.3	5.1	1.3	3.8	11.8	9.8	5.8
SO_4	5.7	6.8	5.8	11.5	4.4	1.6	2.3	5.5
H_2O at 180°C	12.5	15.8	14.8	--	10.2	8.7	11.7	14.7
Compound Composition								
$Al(OH)_3$	11.1	13.7	8.9	23.1	8.2	16.0	13.2	13.9
$Fe(OH)_3$	37.6	34.0	44.3	46.6	24.9	14.2	25.9	44.7
$CaCO_3$	11.4	7.5	7.0	0.0	38.3	25.0	14.4	7.2
$MgCO_3$	--	--	--	--	--	21.0	12.0	6.1
$3MgCO_3 \cdot Mg(OH)_2 \cdot 3H_2O$	25.2	16.4	19.4	5.1	14.4	22.4	24.6	15.6
$CaSO_4 \cdot 2H_2O$	10.7	12.7	10.9	20.7	8.4	3.1	4.4	10.4
Total	96.0	84.3	90.5	95.5	94.2	101.7	94.5	97.9

mine drainage sludges generated from different waters and neutralization processes (3). Since most mine drainages can be classified by broad groupings according to similar chemical composition, the data sufficiently represent the expected chemical properties of mine drainage sludge. The general consensus among researchers is that the chemical makeup of raw water, along with the unit operations and type of neutralizing agent employed, greatly influence sludge formation and associated chemical and physical characteristics.

7.2.2 Physical Properties

Among the many physical properties of mine drainage sludge, the most important are settleability, density, viscosity (sludge flowability), particle size and surface properties, and dewaterability (3).

7.2.2.1 Settleability

Sludge settleability ranks as the most important property with respect to design. Sludges that settle poorly require specialized treatment (i.e., polymer addition) and greatly influence process design. The various alkalis produce sludges with differing properties.

Hydrated lime ($Ca(OH)_2$) is the most popular neutralizing agent in treating mine drainage. Sludges generated from such facilities usually settle well, but compact poorly, producing high volumes of sludge. High sludge volumes have also been reported for quicklime (CaO) treatment. Bisceglia stated that most quicklime products contain 2%-3% unburned limestone or "core." Assuming an 18.1 kkg/d (20 ton/d) usage, this amounts to 364-545 kg (800-1,200 lb) of extra solids (dry) (4). This could represent about 19.0 m^3 (5,000 gal) or more of sludge per day. Also, the compressive settling (compaction) of quicklime sludge is just slightly greater than that of hydrate sludge.

Limestone produces a sludge with good settleability and compaction, but is limited because of pH (7.4 maximum) and the chemistry of iron oxidation (5).

The sodium neutralizing agents (sodium carbonate, sodium hydroxide) produce sludges that are more fluffy and voluminous than the limes. Because they are completely soluble, there is virtually no waste. On a dry solids basis, less sludge is actually produced.

7.2.2.2 Density

Sludge densities are generally reported as percent solids by weight. They are an extremely important parameter in both the initial settling process (pond or clarifier) and final disposal. Densities can vary tremendously, from 0.5% for in situ sludge to 63% for a cake from a dewatering process (6).

The designer can expect a sludge density from settling ponds with a continuous discharge to be anywhere between 0.5% and 4.5%. Clarifier underflows have sludge densities ranging from 1.0% to 7.0%, slightly higher than earthen ponds. Clarifiers offer greater ease in solids handling.

7.2.2.3 Viscosity (Sludge Flowability)

Little has been reported in the literature pertaining to sludge viscosity because the percent solids content is usually so low (less than 5%) that it is readily comparable to water. Nevertheless, sludge flowability is not always that easy. The ability of a sludge to flow is considerably important, especially when designing a collection system or cleaning a pond. Lovell outlines some of the properties and flowability problems encountered when transferring different sludges (3).

1. One sludge moved satisfactorily from the bottom channels of the settling pond to the sludge drying basin by gravity prior to its gelation, although it tended to resist flow from the sloped bottoms to the channel. The limestone-produced sludge was most difficult to move because of its density. Low-velocity flows are preferred during transfer because the tendency toward "ratholing" increases with flow rate.

2. The mobility of an "aged" settled sludge was poor when covered by several feet of water. It would not flow freely from the sloped bottom into the sludge channels and hence through the effluent pipes by gravity flow or from the pump. This fact indicates that the angle of repose of the submerged sludge is greater than 30°; thus, slopes approaching 45° are preferable.

3. Mobility of aged sludge on the sloped surface of a drained pond was satisfactory when movement was initiated by a squeegee or hose water pressure. It tended to move as a large block. On quiescent compression and aging, the sludge forms a gel that has thixotropic tendencies (the property exhibited by certain gels of liquefying when stirred or shaken, then returning to a semisolid form upon standing). The gelation behavior develops within 48 hours under sludge and water pressure compression. There was no evidence of this gelation phenomenon in the settled sludge in the thickener. Apparently, the slow movement of the sludge by the thickener rake prevented gelation. An understanding of aging and gel formation is needed to assist with lagoon design and operation.

   A 0.61- and 0.9-m (2- and 3-ft) thick layer of gelled sludge in one section of a pond was moved into the sludge channel by two men in 4-5 hours using squeegees and low-pressure hoses. Blocks of sludge were cut with shovels; these would slide down the sloping bottom to the sludge channels, provided the bottom was kept thoroughly wet by a small flow of plant water from a hose. Such settled sludge would normally range between 5% and 15% solids. This procedure was rapid and minimized sludge dilution. There was no evidence that the bituminous (asphalt) bottoms significantly enhanced sludge movement. But it did stabilize the pond and eliminate erosion and mud formation problems during sludge removal.

The sidewalls of the pond were earthen. Although no significant erosion problems developed with the compacted clay walls, sludge removal from these areas was more difficult. An occasional small stone from the pond walls would block the check valve of the sludge pump. Bituminous coatings of side and bottom surfaces provide definite advantages and are recommended. Riprap should never be employed below water levels in a pond.

4. Transfer of gelled sludge from the settling pond to drying beds by gravity flow was unsatisfactory. Sludge tends to rathole quickly, allowing excessive amounts of clarified supernatant to move with the sludge and complicate subsequent operations.

5. The sludge had adequate flow characteristics in lines and through pumps when movement began. The sludge transfer lines were flushed with clear water after sludge transfer ended to prevent scaling or further gelation. No line blockage was experienced.

6. Sludges produced with hydroxide-type alkalis were the least dense and did not adhere very tightly to the pond bottom. By contrast, settled limestone slurries were dense, sticky, and claylike in character, and adhered tenaciously to the bottom surface.

The experience and properties outlined above by Lovell alert the designer or operator to the importance of providing for sludge handling and giving forethought to design.

7.2.2.4 Particle Size and Surface Properties

Both of these properties influence the flocculation of sludge particles. The ability of sludge flocs to agglomerate affects the sludge settling velocities, and ultimately, the settling basin size.

The surface property of individual flocs generally refers to the electrostatic charge on the particle. This charge, which causes the particle to resist flocculation, can be offset with the addition of coagulants or polymers.

Researchers have conducted numerous studies on the various types of polymers (3, 7). They found that nonionic (neutral charge) and anionic (negatively charged) polymers were most responsive in AMD sludge treatment. The best flocculant and optimum dosage rate must be established experimentally. Ideally, polymer dosages should range between 0.5 and 2.0 mg/l; however, higher rates have been used.

7.2.2.5 Dewaterability

The ease of removing water from mine drainage sludge is referred to as its dewaterability. Others have defined it as the ease with which sludge can be concentrated into a more manageable form (3). Obviously, the objective of any sludge dewatering operation is to reduce moisture, thereby increasing the solids content, which will minimize disposal volumes. Some basic factors that influence sludge dewaterability are presented in Table 7-2 (6).

TABLE 7-2

SLUDGE DEWATERABILITY VARIABLES

1. Initial concentration of solids
2. Age and temperature
3. Viscosity
4. Compressibility
5. Chemical composition
6. Physical characteristics

These variables apply mostly to mechanical methods of dewatering. In most cases mine drainage sludge undergoes gravity dewatering or air drying either by lagooning or disposal into deep mines.

7.3 Methods of Mine Drainage Sludge Dewatering and Disposal

This section describes sludge dewatering and disposal methods and includes lagooning, deep mines, underground disposal guidelines, filtration, bed drying, and centrifugation.

7.3.1 Lagooning

Lagoons offer the easiest and one of the cheapest methods of sludge storage and dewatering when land is available. Mine drainage treatment plants are usually located in remote or isolated areas where land is readily available.

Lagooning can serve the following three purposes in mine drainage treatment: (1) as settling ponds or impoundments (Chapter 6) where sludge is collected and undergoes preliminary dewatering; (2) as the primary sludge dewatering unit; or (3) as permanent storage or disposal facilities.

Lagooning has been used in every mode of operation (singular, series, parallel) to enhance sludge consolidation and thickening. Large, single lagoons (settling impoundments) perform all three functions the least effectively. Very little dewatering and compaction of the sludge takes place in these facilities. The continuous presence of water above the sludge zone does not allow any atmospheric dewatering. The designer should not consider a large, single lagoon (impoundment) as an effective sludge dewatering facility, although it may have certain advantages as a settling unit, as discussed in Chapter 6.

Settling ponds in series, where the first serves as the primary settling unit and the second as a polishing pond, are often used in mine drainage treatment. This system offers better sludge control by isolating the majority of the solids in the first pond. This arrangement has the same disadvantage as a large single pond where little sludge dewatering occurs. In this arrangement, the primary settling pond is usually equipped with a sludge removal device, and the sludge is transferred to a disposal lagoon where atmospheric dewatering can occur.

A dual or parallel arrangement of settling ponds, together with an isolated dewatering lagoon, may be the optimum system for achieving the most effective sludge dewatering possible through natural methods. Where land is available, construction of two settling ponds, each with sufficient volume to treat the design flow, appears ideal. This system allows alternate use of the ponds so the inactive pond can undergo first-stage dewatering. Its contents are then transferred to the final disposal lagoon for further dewatering. In both ponds, supernatant or surface water must be decanted to allow the sludge full exposure for natural drying.

Holland et al. experienced good results with this method of drying where a raw sludge with an initial solids content of near 1% dewatered to 14% within 3 weeks. They further observed that solids can be increased by 8% and can dewater as much as 20% solids. These are remarkable results and should not be assumed to be easily achieved.

After most of the water is decanted from a lagoon, shrinkage cracks commonly occur that will honeycomb the entire pond. A unique property of dewatered sludge, if not totally submersed, is that it will resist "rewetting," which enhances the dewatering process.

The problem of ultimate sludge disposal still remains. Closure of sludge disposal lagoons presents a formidable problem. Sludge lagoons are very difficult to cover with soil. In such an operation, extreme care should be taken as the sludge exhibits thixotropic tendencies; i.e., it will tend to liquefy upon vibration. Dried sludge samples have exhibited relatively low unconfined compressive strengths in the range of 200-400 lb/ft^2. Such a sludge would be under stress during covering with soil, especially by the weight and movement of the machinery. Special safety precautions are needed for operators of the covering equipment.

Spread burial of dried sludge can be more economical if it can be done "in-house" and the land area for spreading is available. This method takes advantage of the soil conditioning benefits of the sludge (i.e., alkalinity, minerals). Conrad, in a test using sludge to treat a corn field, reported that the corn grew faster and taller and produced a better yield than the control plot (8). Controlled amounts of sludge applied to an area, then covered by soil in layers, can be utilized as a means of sludge disposal and improving the land. The heavy metal content and other toxic constituents of the sludge should be evaluated for their long-term effect on the vegetation. Sludge should not be applied to high-acidity soils (i.e., gob piles) where the alkalinity will totally leach out and the metals will redissolve.

In summary, lagoon drying can substantially reduce sludge volumes when operated properly. The best results are obtained when undisturbed dewatering is allowed to occur after free water is decanted. The designer should remember that deep sludge lagoons may not be satisfactory for ultimate disposal.

7.3.2 Abandoned Deep Mine Disposal

Disposal of mine drainage sludge to abandoned deep mines by means of a borehole is exploited whenever possible. The apparent simplicity of operation is not always indicative of the practice. As with any disposal method, all the legal and environmental ramifications must be considered before it becomes possible. Nevertheless, this means of disposal offers the most economical and simplest method for sludge disposal.

Steinman, because of restricting terrain at a treatment site, could not construct a lagoon of adequate size for sludge disposal (9). It was decided that sludge disposal would be to an abandoned deep mine 14.5 km (9 mi) away. Even under these conditions where extra handling and trucking were required, it was proven economical and successful, although a thorough inspection of the proposed underground area by the state regulatory agency was required before permitting the site.

An example of possible state agency guidelines for underground disposal follows.

7.3.2.1 Guidelines for Underground Disposal of Sludge from Acid Mine Drainage Treatment (10)

Whenever underground disposal of sludge from treatment of mine drainage is proposed, the sludge must have a pH of 7.0 or above, and all of the iron must be in the ferric form.

Any application submitted that proposes underground disposal of sludge should contain the following supplemental information:

1. Location
    a. Name of the abandoned mine in which the sludge is to be disposed.
    b. Outline of the mine workings on the latest available U.S. Geological Survey topographic map.
    c. Location of the discharge point of the sludge shown on the latest topographic map.

2. Mine Hydrology
    a. Is there water in the mine? If water is present, what area of the mine is flooded?
    b. Is the water level in the mine rising, falling, or static?
    c. Does the flooded mine discharge? If so, supply the following:
        (1) Location of discharge point(s).
        (2) Quantity of the discharge(s).

    - (3) Elevation of the discharge(s).
    - (4) What quality controls will be at these discharge points to determine future variations in quantity and quality?
  - d. What is the extent of previous mining next to and below the proposed disposal area?
  - e. Does the disposal mine have any connections with other nearby workings?
    - (1) List discharges to other mines.
    - (2) List discharges from other mines.
  - f. Is there pumping from the disposal mine or any interconnected mines? If so, supply information concerning quantity and quality of pumpage.

3. Quality of Mine Water

  - a. Supply the following information concerning the quality of water in the mine where the sludge will be disposed:
    - (1) pH.
    - (2) Acidity or alkalinity.
    - (3) Iron.

4. Input of Sludge

  - a. Volume of sludge.
  - b. Volume in the mine available for sludge disposal.
  - c. Estimated length of time sludge will be disposed of in the mine.
  - d. Leachate quality from the sludge (ASTM-A Method).
  - e. The concentration of sludge.
  - f. What effect will the sludge have on any discharges from the mine?
    - (1) Quantity.
    - (2) Quality.
  - g. Will any of this sludge be flushed out by movement of the impounded mine water?

5. Geology

  - a. Structure of the coal beds.
    - (1) Strike.
    - (2) Dip.
  - b. Are there any faults present in the immediate area?
    - (1) Location.
    - (2) Strike and dip.
  - c. What is the elevation of the local groundwater table? How was this determined?

6. When the pH of the water in the abandoned mine is above 4.0, the ferric iron is basically insoluble. At a pH below 4.0, ferric iron is soluble. When it is proposed to dispose of iron sludge into water with a pH below 4.0, supporting data must be supplied to show that the proposed disposal will in no way adversely affect any present or future discharges from the mine pool.

Little detailed information can be found in the literature pertaining to deep mine disposal of drainage sludge. The most common process is to withdraw the sludge from the settling basin by means of a portable or permanent collection system and pump it directly to the borehole. No dewatering is considered before injection. Therefore, the solids content is low (0.5%-2.0%) and high volumes must be handled.

7.3.3 Vacuum Filtration

Vacuum filtration, which has long been employed in sewage treatment for sludge dewatering, is readily applicable to mine drainage sludge. The principle of operation is simple. The revolving drum is perhaps the most common type of vacuum filter. It has a series of vacuum cells that run the length of the drum. The drum, turning at less than 1 r/min, passes through the sludge where a vacuum of 0.40-0.87 atm (12-26 in of mercury) is applied to the submerged portion (approximately 25% of the periphery), drawing a cake to the filter media surface. The filter media, referred to as a "cloth," can be made from a number of materials, the most common of which is polyethylene. As the sludge emerges from the reservoir, the vacuum dewaters the cake as it rotates on the drum. The cake is then removed by a scraper, a blast of air, or coiled springs.

Some of the operational variables that influence the dewaterability of the sludge by this method are listed in Table 7-3 (6).

TABLE 7-3

VACUUM FILTRATION OPERATIONAL VARIABLES

1. Amount of vacuum
2. Amount of drum submergence
3. Drum speed
4. Filter media
5. Sludge conditioning before filtration

Lovell, operating a vacuum filter with the cycle times listed in Table 7-4, was able to achieve the filtration rates presented in Table 7-5 (3).

Lovell states that filter feed slurries with less than 2% solids consistently produced the lowest solids filtration rate, 73.9-147.8 kg/m^2/d (15-30 lb/ft^2/d). Solids at 5% or greater, however, can have filtration rates of 610-1,616 kg/m^2/d (125-331 lb/ft^2/d), indicating that vacuum filtration can be economically feasible (3). Other researchers studying vacuum filtration reported successful results in producing a filter cake with 24%-35% solids (7, 11). At this percent solids, the sludge can be easily handled and is acceptable for landfill disposal.

TABLE 7-4

OPERATIONAL CYCLES

Submergence	16 s	21%
Drying time	50 s	64%
Discharge time	12 s	15%
Total cycle time	78 s	100%

Cycles/hr - 46.1

Corresponding filter drum speed - 0.77 r/min

7.3.4 Pressure Filtration

Pressure filtration is merely an acceleration of the vacuum filtration process. Based upon the same principle (a pressure differential across a filter media), pressure filters can consist of plates or of plates and frames. Sludge enters the filtration chambers, designed so that liquid (filtrate) passes through the filter medium while the solids (cake) are held within the chamber. Sludge does not flow from chamber to chamber (series), but enters each chamber independently (parallel) so that each area fills with solids at the same rate, retaining the same quantity of filter cake and passing nearly identical filtrates. The filtration process continues until the chamber area is full or a predetermined terminal pressure is reached, completing the cycle. The sludge feed is stopped and the chambers are opened to remove the filter cake. The cycle is then repeated. Figure 7-1 illustrates a simple manually operated filter press.

The rate of sludge filtration depends upon (1) feed pressure, (2) thickness of filter cake, (3) sludge temperature and viscosity, (4) nature of the cake solids, and (5) the filter medium.

Pressure filtration of mine drainage sludge or other waste sludges has not been widely employed in the United States because of high labor, maintenance, and capital costs. Rummel, in East Germany, investigated pressure filtration of mine drainage sludge, and found that an influent slurry of 1.2% solids could only be dewatered to a filter cake of 20%-30% solids after filtration (12). The best filtration rate reported was 50 $l/m^2/hr$ (0.061 $gal/m/ft^2$). Unfortunately, this particular treatment plant generated 40,000 m^3 (10,000 gal) of sludge per day. Thus, it was concluded that the filter output was insufficient for the application.

Several sludges were dewatered by this method with moderate success. Table 7-6 presents the results of experiments conducted by Akers et al. (13). The data indicate that good filtration rates could be obtained without floccu-

TABLE 7-5

MINE WATER NEUTRALIZED SLUDGE SOLIDS FILTRATION RATES

	Sludge Filtration Rates, lb dry solids/ft^2/24 hr				
Mine Water Source	% Solids in Filter Feed Slurry	Fresh Hydrated Lime	Old Hydrated Lime[a]	Pebble Dolomitic Lime	Dolomitic Hydrated Lime
Proctor No. 1	1.62	26.4[b] 27.8			
Proctor No. 1	2.53			54.0	
Proctor No. 2	2.70	62.7 72.0			53.2 53.8
Proctor No. 2	7.19		131.0 137.0	63.0 64.0	
Proctor No. 2 (bio)	1.48		39.6 50.3		
Proctor No. 2 (bio)	3.26		82.5 81.4		
Proctor No. 2 (bio)	7.18		331.0		
Bennett's branch	1.96	33.1 38.6[c]			
Bennett's branch	2.72	40.1 42.9			
Tyler run	1.58			15.6 21.2	
Tyler run	5.03			128.5 125.3	

---

[a] In this case, hydrated lime, stored in the plant lime storage silo for a period of 3 months, converted to about 50% $CaCO_3$.

[b] The double values reported for most tests represent duplicate determinations made on each sludge slurry.

[c] No precoat material employed.

lants. Also, it was concluded that pressures lower than 5.1 atm (60 lb/in^2g) should not be used, because the principal advantage of this dewatering method is a high filtration rate per unit of filter area requiring high pressure.

The capital and operating costs for a full-scale process are shown in Table 7-7 for a plant being built in 1972. Since then many of these prices have doubled, thus adding to the unattractiveness of this method.

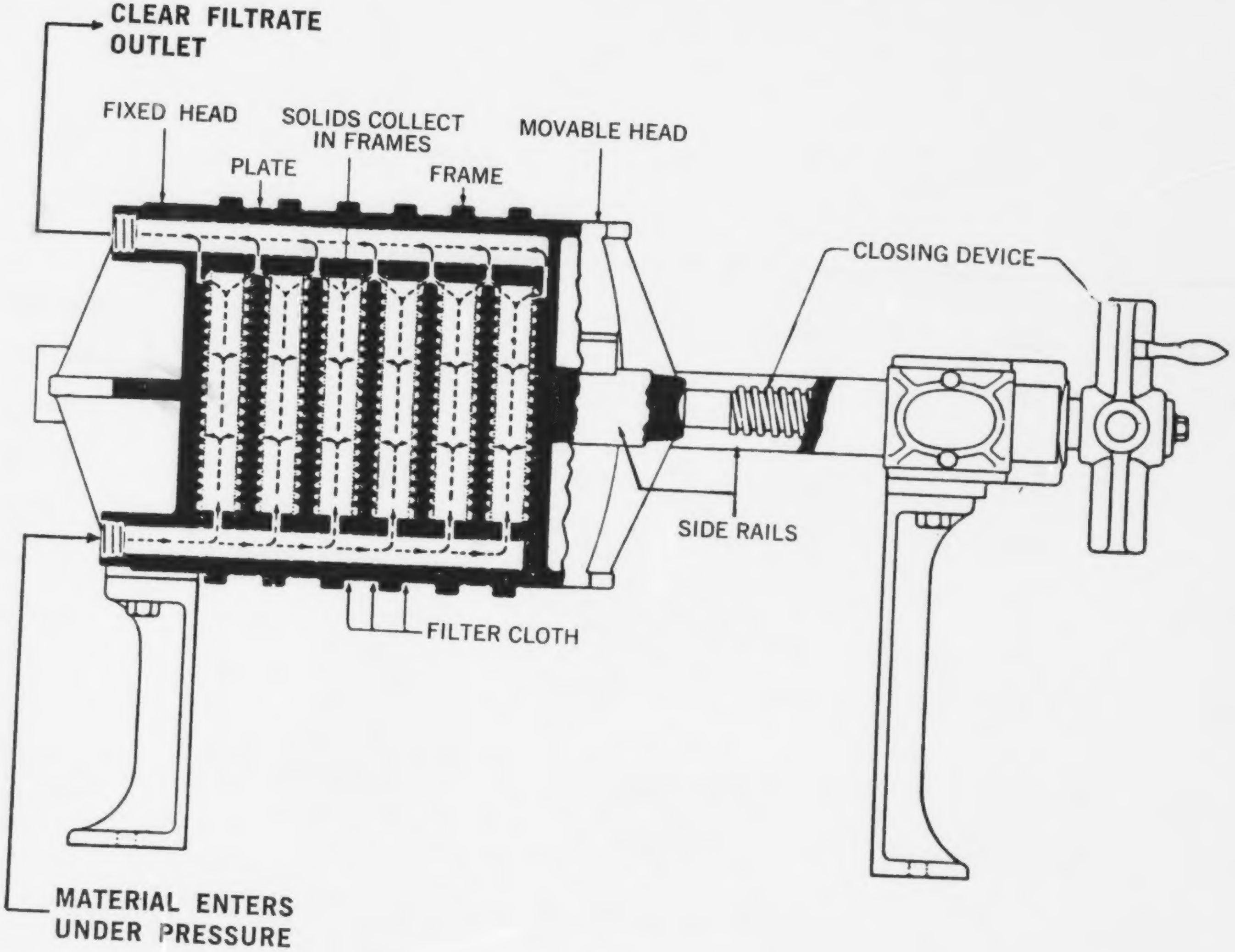
CLEAR FILTRATE OUTLET
FIXED HEAD
PLATE
SOLIDS COLLECT IN FRAMES
FRAME
MOVABLE HEAD
CLOSING DEVICE
SIDE RAILS
FILTER CLOTH
MATERIAL ENTERS UNDER PRESSURE

Figure 7-1. Manually operated filter press.

TABLE 7-6

PRESSURE FILTRATION CAKE DATA

Sludge Used	Pressure		Thickness of Final Cake		Filtration Time to Produce Cake	Final Solids of Cake
	atm	lb/in^{2}g	cm	in	min	%
Norton	5.083	60	19.0500	7.5000	179	20.8
Edgell	5.083	60	1.2700	0.5000	198	26.2
Edgell	6.444	80	0.9525	0.3750	138	26.0
Edgell	7.805	100	1.4288	0.5625	175	26.0
Banning	7.805	100	2.2225	0.8750	219	11.8
Shannopin	5.083	60	6.0325	2.3750	170	8.7
Shannopin	6.4444	80	6.9850	2.7500	180	12.0
Shannopin	7.805	100	7.3025	2.8750	200	10.9

All sludges were air-blown for 5 minutes after break.

TABLE 7-7

PRESSURE FILTRATION - NORTON TREATMENT PLANT SLUDGE (PRELIMINARY PRICES - SPRING 1972)

Equipment	Capital Costs	Costs/yr	Costs/d
1 48-in filter press	$21,000.00		
1 plate shifter	2,700.00		
Feed pump with accessories	900.00		
Precoat equipment	5,000.00		
Construction and installation: 35% of equipment	10,000.00		
Total equipment cost	$40,000.00		
Equipment depreciation		$4,000.00	$ 11.00
Building: 1,200 ft^2 at $10.00/$ft^2$	12,000.00		
Building depreciation		$ 400.00	1.00
Total capital cost	$52,000.00		
Operational costs			
Maintenance: 6% of total capital cost		$3,120.00	8.50
Electricity: 270 kW-hr/d at $0.0175 kW-hr			4.70
Labor: 2 men at 24 hr at $6.00/hr each			$288.00
Precoat: Johns-Manville Celite 501, $73.00/ton F.O.B. Ca. warehouse, $105.40/ton delivered Morgantown, W. Va.			50.00
Total capital and operational cost			$363.20

Assuming 50,000 gallons clarifier underflow/d: $7.30/1,000 gal sludge dewatered

Assuming 20,000,000 gal acid water/d: $0.02/1,000 gal acid water

7.3.5 Porous Bed Drying

Sludge drying beds constructed of graded materials have long been employed as a method of dewatering. A variety of filter media, such as sand, crushed limestone, coal, red dog, and gravel, have been used. Water is removed in this dewatering method by decanting the ponded surface water, by percolation through the bottom of the bed, and by evaporation.

Figures 7-2 and 7-3 illustrate the drying bed constructed by Grube and Wilmoth in their experiments to evaluate this method of sludge dewatering (14). Modeled after drying beds traditionally used for sewage sludge, it is obvious that detailed construction is required, although others have employed beds with less material classification. For example, just two layers of materials, crushed limestone beneath coarse sand, can be used, with a minimum depth of 15.24 cm (6 in) for each medium.

Lovell summarizes the sludge drying bed area requirements in the design graph shown in Figure 7-4.

Based on data from studies by Lovell, a 3,785 m^3/d (1 Mgal/d) plant requires 6,691 m^2 (72,000 ft^2) of bed area. For all practical purposes, this method of sludge dewatering could only be feasible for a plant with low sludge production. Even then, the climate of the northeastern United States does not favor this method.

Grube and Wilmoth described the operational difficulties encountered with the application and removal of sludge from drying beds (14). They concluded that drying bed operations that allowed sludge freezing rendered this method of dewatering unacceptable.

Summer operation with lime-neutralized, coagulant-treated sludge provided the following observations:

1. Approximately 20% by volume of the influent sludge (at 2.5% solids) was retained on the sand bed as a 20% solids gellike mass, while 50% drained through the sand as a clear effluent. The remaining 30% was assumed to be lost to the atmosphere by evaporation.

2. The drainage rate through the sludge and sand averaged 26 l/d/m^2 (0.6 gal/d/ft^2).

3. Sludge solids appeared to stabilize near 20% within 20 days drying time.

Although this method of sludge dewatering will eventually produce a cake, removal from the bed remains a problem. Grube and Wilmoth attempted to remove the sludge with a rubber-tire highlift, but this proved almost impossible and required rebuilding portions of the bed (14). The recommended alternative was a trac-mounted loader, which functioned well. The large volumes of sludge and weather conditions that occur in the areas where mine drainage treatment plants are most prevalent (eastern United States) make this method of sludge dewatering impractical.

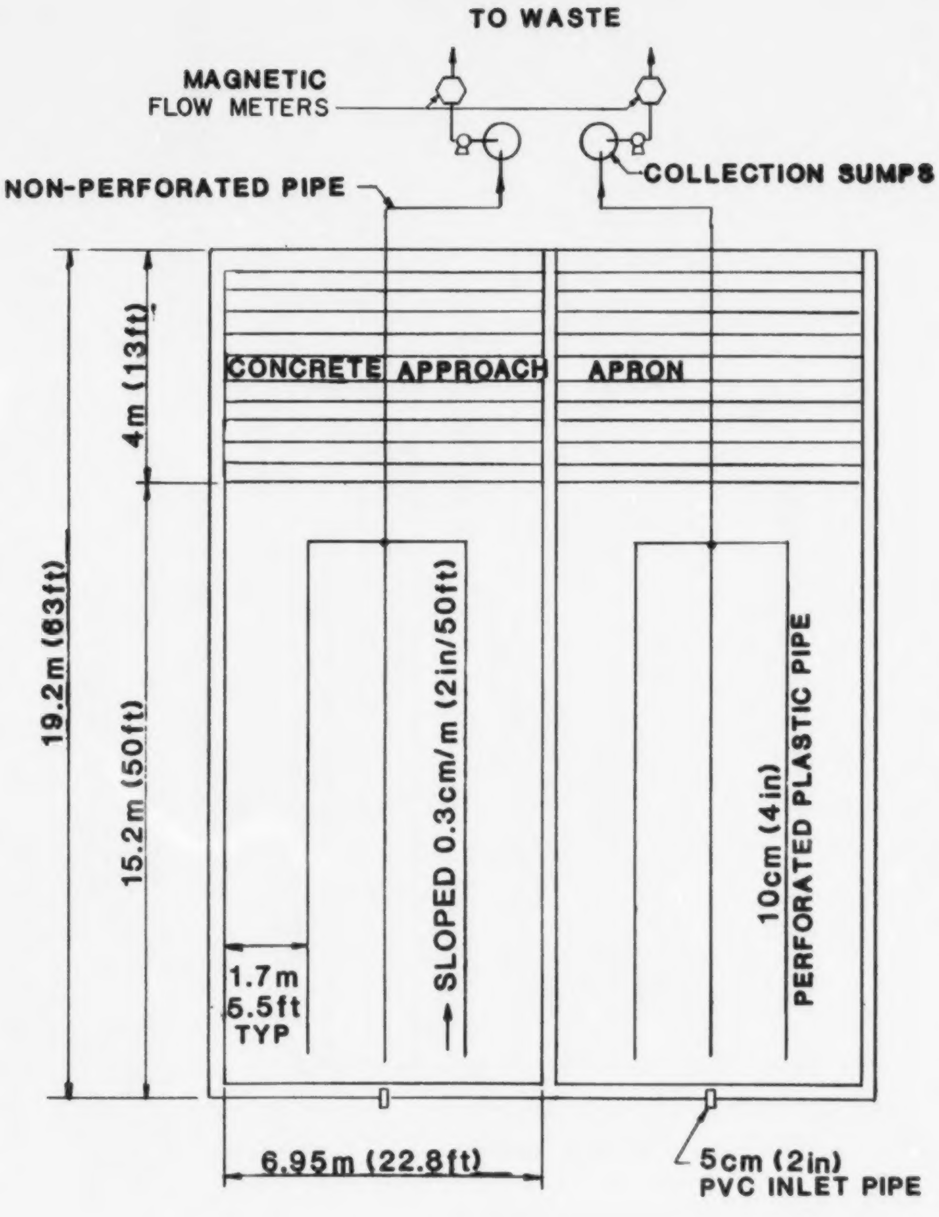
TO WASTE
MAGNETIC FLOW METERS
NON-PERFORATED PIPE
COLLECTION SUMPS
CONCRETE APPROACH APRON
4m (13ft)
19.2m (63ft)
15.2m (50ft)
SLOPED 0.3cm/m (2in/50ft)
10cm (4in) PERFORATED PLASTIC PIPE
1.7m 5.5ft TYP
6.95m (22.8ft)
5cm (2in) PVC INLET PIPE

Figure 7-2. Top view of drying bed construction.

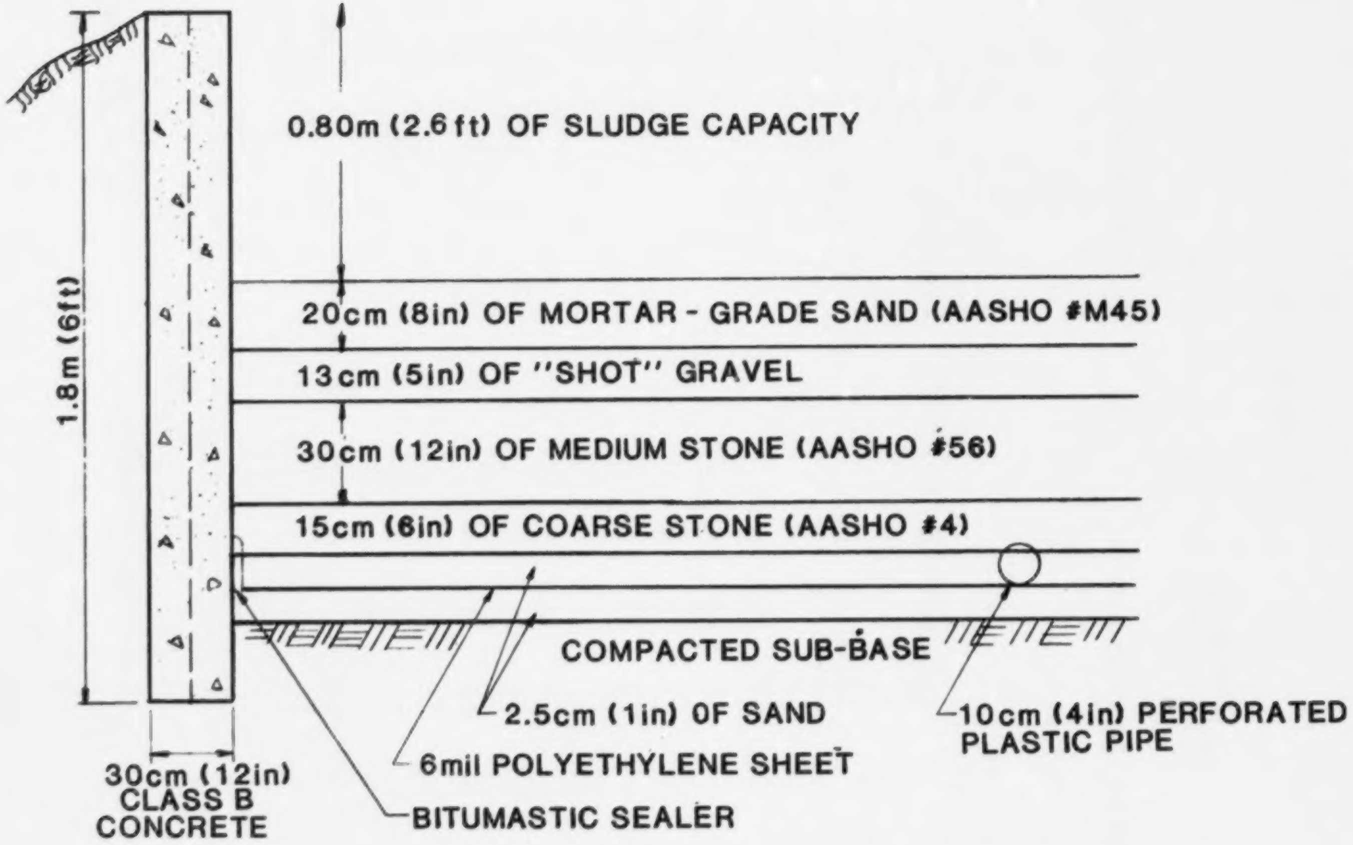
0.80m (2.6 ft) OF SLUDGE CAPACITY
20cm (8in) OF MORTAR - GRADE SAND (AASHO #M45)
13cm (5in) OF "SHOT" GRAVEL
30cm (12in) OF MEDIUM STONE (AASHO #56)
15cm (6in) OF COARSE STONE (AASHO #4)
COMPACTED SUB-BASE
2.5cm (1in) OF SAND
6mil POLYETHYLENE SHEET
BITUMASTIC SEALER
10cm (4in) PERFORATED PLASTIC PIPE
1.8m (6ft)
30cm (12in) CLASS B CONCRETE

Figure 7-3. Cross-sectional view of the drying bed construction.

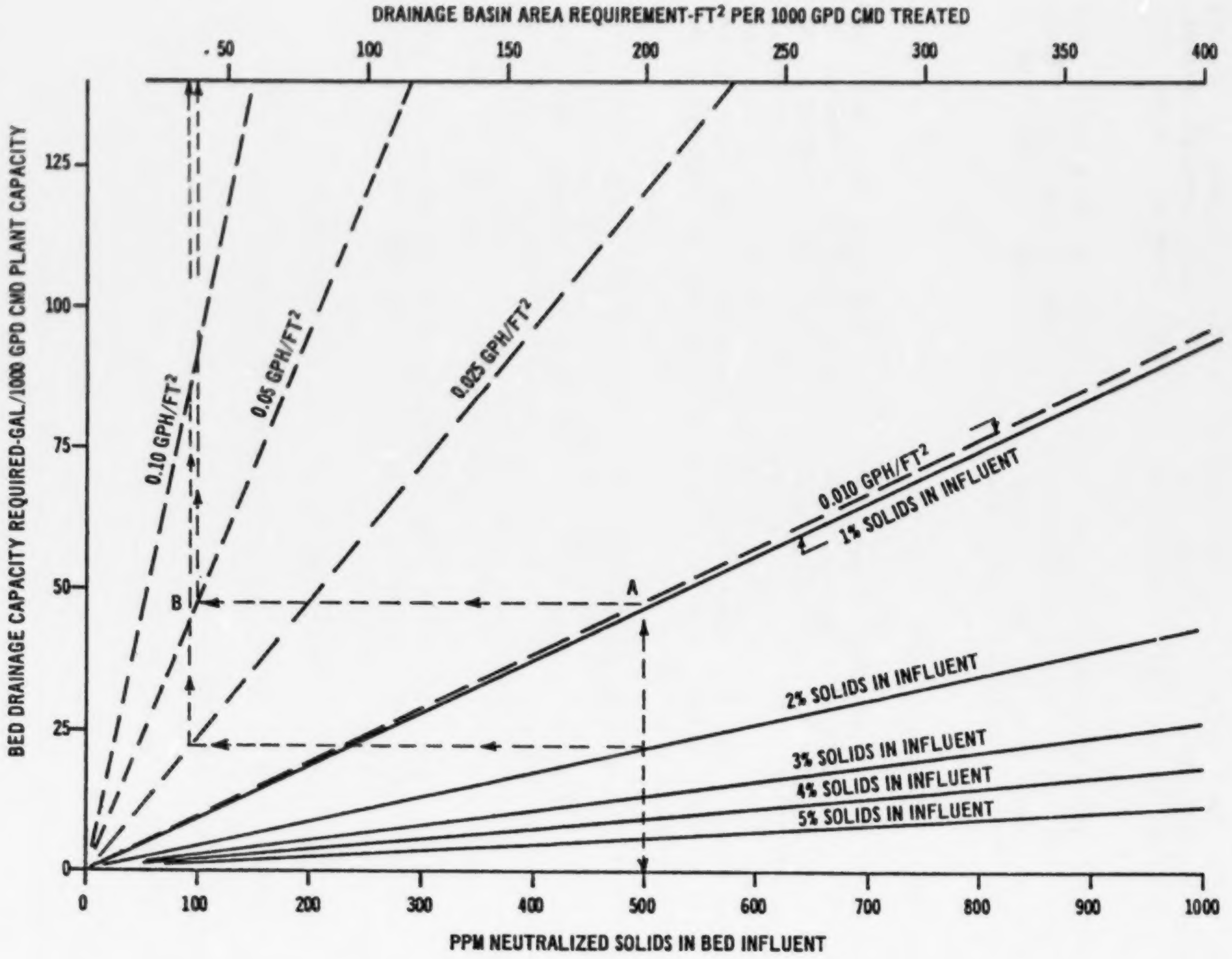

Figure 7-4. Sludge drying bed sizing requirements.

7.3.6 Centrifugation

Centrifugation is the separation of substances of different densities by the use of centrifugal force. Many types of centrifuges are manufactured, but all incorporate the following three basic operations: (1) a feed system which delivers the sludge; (2) a revolving solid bowl or basket to collect dewatering sludge; and (3) a sludge and skim (effluent) removal system. Dewatered sludge can be removed either by a scraper blade or a screw conveyor.

Akers et al. conducted experiments on several sludges using a solid bowl centrifuge. Table 7-8 summarizes results of the sludges tested (13). These results vary drastically for each sludge, indicating that a centrifuge is effective for dewatering some, but not all, sludges.

7.4 High-Density Sludge Process

Researchers from the Bethlehem Steel Corporation Research Department and Bethlehem Mines Corporation cooperated in the development of a process designed to provide both improved settling characteristics and increased sludge solids concentration. The result of their research is the high-density sludge process (HDS). This process reportedly achieves settled sludge concentrations between 15% and 40%, compared to a maximum of 15% from conventional lime neutralization. The resulting sludge storage or disposal volume is thus reduced by a significant factor.

The major differences between the HDS and the conventional lime neutralization processes are shown in Figures 2-2 and 7-5. The HDS process has a second reaction tank for mixing return sludge and lime slurry. All other unit processes are the same as conventional lime neutralization. A clarifier is necessary to provide adequate process control of sludge settling and thickening and to provide an efficient means of sludge collection and return.

Research conducted on both pilot plant and full-scale operations has pinpointed several operating parameters that affect the HDS process. These parameters are as follows (15):

1. ferrous-to-ferric iron ratio in the acid mine drainage;
2. recirculated solids-to-precipitated solids ratio;
3. point of alkalinity addition;
4. neutralization pH;
5. reaction tank detention time.

To elaborate, the following discussion of each is offered.

TABLE 7-8

SUMMARY OF CENTRIFUGATION TEST*

Feed Solids (5)	Flow Rate l/sec	Flow Rate (gal/min)	Bowl Solids %	Skim Solids %
Shannopin sludge				
2.27	0.03	(0.5)	33.4	11.0
2.08	0.06	(1.0)	33.6	8.8
2.09	0.09	(1.5)	32.5	9.0
2.04	0.12	(2.0)	36.9	8.0
2.24	0.15	(2.5)	35.4	7.7
1.85	0.03	(0.5)	26.6	13.8
Banning sludge				
0.64	0.03	(0.5)	8.8	5.1
0.68	0.09	(1.5)	8.4	3.8
0.69	0.12	(2.0)	8.7	4.6
0.72	0.15	(2.5)	8.6	5.0
0.73	0.18	(3.0)	8.1	5.6
0.99	0.03	(0.5)	11.4	9.2
Norton sludge				
1.54	0.03	(0.5)	41.3	12.9
5.52	0.03	(0.5)	53.7	19.2
10.05	0.03	(0.5)	64.1	17.5
3.77	0.03	(0.5)	63.0	22.1
2.09	0.06	(1.0)	44.9	11.6
2.48	0.09	(1.5)	50.5	13.5
2.73	0.12	(2.0)	55.8	12.8
6.12	0.06	(1.0)	51.8	19.2

---

*Capital and operating costs ranged between $1.80 and $4.50 per 2.785 m^3 (1,000 gal) of treated sludge based upon 1972 prices.

7.4.1 Ferrous-to-Ferric Iron Ratio

The percentage of ferrous iron in the acid mine drainage has a significant effect on the maximum settled solids concentration that can be attained by the HDS process. Test data have been used to generate a curve representing the relationship between the sludge density and the ferrous iron percentage in the influent (Figure 7-6) (2). This curve shows that ferrous iron percentages below 70% had relatively little effect on the concentration of

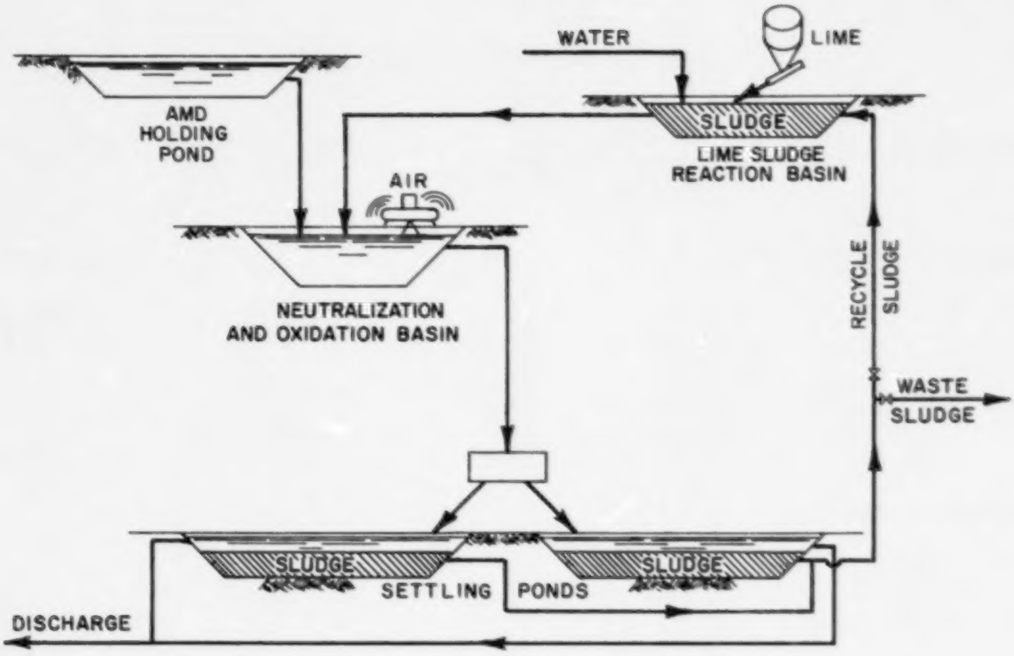
WATER
LIME
AMD HOLDING POND
SLUDGE
LIME SLUDGE REACTION BASIN
AIR
NEUTRALIZATION AND OXIDATION BASIN
RECYCLE SLUDGE
WASTE SLUDGE
SLUDGE
SLUDGE
SETTLING PONDS
DISCHARGE

Figure 7-5. High-density sludge neutralization process.

sludge solids produced. For ferrous iron percentages greater than 70%, the sludge concentration approached 50%. This highest sludge solids concentration was achieved under laboratory conditions with synthetic mine drainage.

7.4.2 Recirculated Solids-to-Precipitated Solids Ratio

The ratio of recirculated solids to precipitated solids has a direct effect on the settled sludge density. The trend is for the sludge density to increase as the ratio of recirculated to precipated solids is increased. This increase in rapid up to a recirculation ratio of 20:1, moderate between 20:1 and 30:1, and small above 30:1. The recommended optimum recirculation range is 25:1 to 30:1 (15). Operation within this range maximizes sludge density while minimizing the clarifier area requirement.

7.4.3 Point of Alkalinity Addition

The HDS process operates successfully only when the lime slurry and recycle sludge are mixed in a reaction tank prior to the addition of acid mine drainage and aeration. This represents the most critical step in the process. Any other process arrangement results in the failure of the process to

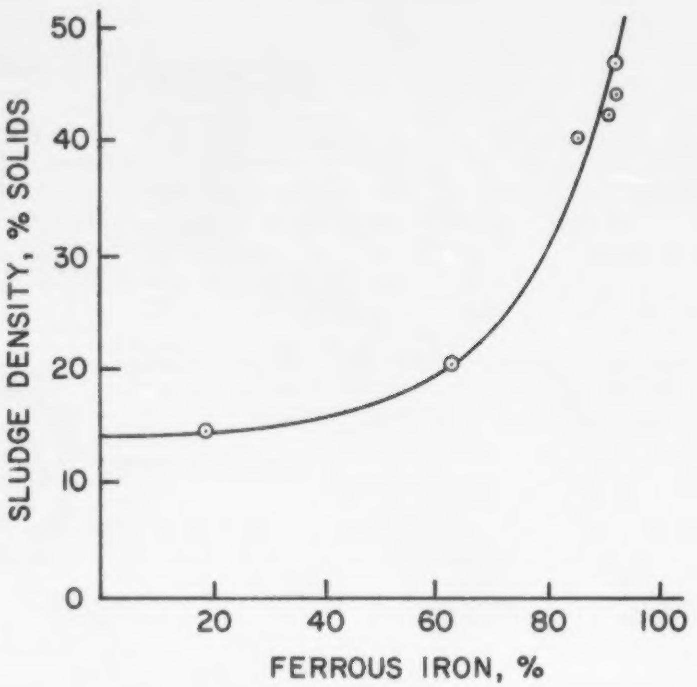

Figure 7-6. Sludge density vs. ferrous iron percentage for the HDS process.

achieve the desired solids concentration.

7.4.4 Neutralization pH

The neutralization pH affects the settled sludge concentration, the oxidation rate, and the settling area requirements. The optimum operating pH was found to be within the 7.2-7.7 range for their AMD (16). Within this range, maximum sludge densities are produced, a satisfactory iron oxidation rate is maintained, and mechanical clarifier area requirements are minimized. Operation within a pH range of 6.0-6.5 results in reduced lime usage, but the ferrous iron oxidation rate is unacceptably slow. Operation within the pH range of 8.2-8.7 reduces the sludge solids concentration from 35% to 20% and increases the mechanical clarifier area requirement. Operation within the 9.0-9.5 range produces a rubbery sludge that hinders pump and clarifier operation.

7.4.5 Reaction Tank Detention Time

Separate tanks must be provided for mixing the lime slurry with the return sludge and for the neutralization-oxidation reaction. A 1-minute detention time in the sludge reaction tank is adequate to maintain a sludge of maximum density. The detention time in the neutralization-oxidation tank is a function of the ferrous iron concentration and the operating pH.

The feasibility of this process has been demonstrated by Bethlehem at a test plant near Ebensburg, Pa., and at several full-scale facilities. Sludge concentrations generally conformed to the predictions made by previous small-scale tests (17). The HDS process, however, is best suited to treat high ferrous iron mine drainages. Conventional neutralization process plants with mechanical clarifiers can be easily converted to the HDS process by the addition of a sludge line and a sludge reaction tank.

In summary, the high-density sludge process offers a method for improving the sludge concentration of the settled sludge. Nevertheless, an aura of skepticism exists among designers in the field of mine drainage as to the actual upper limits of performance. Currently, only one publication has reported a sludge density of 10%-12% in a full-scale mine drainage treatment plant.

The Bethlehem Corporation holds a patent on the HDS process; thus, their approval of certain designs is required.

7.5 Summary

The most practical and economical method of sludge disposal is pumping to an abandoned deep mine. The designer should consider this alternative whenever possible, but only after acquiring a permit that addresses the environmental impact of this practice.

Another viable alternative is to thicken the sludge in a lagoon to a form that is more easily handled. The thickened sludge can ultimately be disposed of in a reclamation program for an active stripping operation, or mixed with the refuse tailings from a vacuum filter for a coal preparation plant. Both cases require a good control program to minimize operational problems. A tailings-to-sludge ratio for a good spreadable mix varies with the individual moisture content of the two materials involved. Probably the easiest method for determining a workable tailings or soil-to-sludge mixing ratio is by trial and error. After a short period of time, the disposal operator will become fairly efficient at judging the proper mixing ratio to achieve compaction and burial. Obviously, if a large amount of sludge requires disposal and insufficient quantities of dewatered tailings exist for satisfying a proper ratio, another disposal method must be considered.

Table 7-9 presents a summary of the results and costs for several sludge dewatering methods taken from studies conducted by Akers (13) and the Coal Research Bureau (6).

In conclusion, sludge disposal will remain the most formidable problem facing designers and owners of mine drainage treatment plants. Any method of sludge handling, dewatering, and disposal will incur cost. The most economical method is deep mine disposal. It is important, however, to emphasize that

TABLE 7-9

COAL MINE DRAINAGE DEWATERING METHODS

Dewatering Method	Sludge Cake % Solids	Costs[a] ($/3.875 m³ (1,000 gal dewatered sludge)
Vacuum filtration	8.8 - 30.9	3.40
Precoat vacuum filtration	11.4 - 35.1	1.40 - 3.50
Porous drying beds[b]	15.0 - 25.0	4.80 - 19.10
Pressure filtration	8.7 - 26.2	1.70 - 7.30
Centrifugation	8.1 - 64.1	1.80 - 4.50
Single lagoon	0.5 - 4.5	c
Drying lagoon	12.0 - 20.0	c

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[a] Akers, February 1973

[b] After Lovell and Grube

[c] Assessment based on land value and ultimate disposal (labor)

the environmental acceptability of this method is unknown at this time. In fact, the environmental acceptability of all of the ultimate disposal techniques needs to be assessed.

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CHAPTER 8

ELECTRICAL REQUIREMENTS AND INSTRUMENTATION

8.1 Introduction

The purpose of this chapter is to inform the designer about electrical power, motors, and various instrumentation. Explained herein are some common sense rules in the design and arrangement of equipment, along with suggestions for a clean, efficient operation.

8.2 Electrical Power

The design of equipment for a mine drainage treatment plant depends upon quantity and quality of raw water to be treated. Consequently, process units, mixers, aerators, sludge pumps, and other related equipment requiring electricity must be sized accordingly.

The question of electric power supply is a very individual situation, depending upon the size of the plant and the form in which local electrical power is available.

Three-phase power is almost always available and should always be used to (1) keep conductor size down; (2) allow use of standard motors; (3) keep motor maintenance at a minimum; and (4) provide the most reliable plant operation.

The voltage to be used (230 or 460) is somewhat dependent on what voltage is locally available. In remote areas, high-tension power is the most likely possibility; the local power company will provide the necessary step-down transformers. In populated areas, even though 230- or 460-V (volt) power may be available, the plant load requirements may still make it necessary for the power company to provide a separate step-down transformer off the high-tension feeder. Simply stated, either 230- or 460-V 3-phase power can usually be provided by the local power company. Actually, the problem of fulfilling the electric power requirements for a plant can readily be solved as follows:

1. Determine from the equipment manufacturers the load requirements (hp) of each motor (including the largest to be used in the plant).

2. Use 460-V supply if there are motors larger than 10 hp, since 460-V supply is more economical.

3. Use motors of the same voltage throughout the plant.

4. Always use 3-phase power.

5. After the above requirements are determined, contact the local electric power company; they will decide how they can best provide the service required.

8.3 Motors and Electrical Controls

Every motor requires a degree of control. The control is dependent upon the sophistication required. As a minimum and in accordance with the usual application, each motor is controlled from its own motor control center (MCC) compartment, which contains the motor starter contactor, motor overload protection, and usually a "start-stop" push button and "run" light on the front cover of the individual compartment. It is wise to provide a "lock-out" feature in the MCC to protect maintenance personnel when a motor is out of service.

More sophistication and larger plant controls may require a control panel on which motor controls (usually start and stop) are located remotely from the MCC, as well as fault annunciators. Control cabinets may also be required to house control devices if automatic sequencing is involved.

No matter what degree of motor control is used, the control equipment should be located in the cleanest area possible away from dust, fumes, and moisture. A separate control room is usually the best solution.

The equipment manufacturers will recommend the appropriate housing for the motors. This is directly related to the type of service and the environment the equipment will experience.

8.4 Instrumentation

8.4.1 pH Control Systems

The heart of most mine drainage treatment plants is the pH control system. Generally, it is wired to an interlock system that terminates operation if the pH wanders outside preset limits (these limits are usually pH 6.0 and 9.0).

There are many types of pH electrodes on the market, each with advantages and disadvantages. The designer should look for probes that can withstand daily handling and a certain amount of abuse, and are easy to standardize. The extrasensitive laboratory-type probes should be avoided, along with those directly mounted in pipeline assemblies.

The typical immersion assembly, where the electrodes contact only the neutralized stream and the cable assembly remains emerged, are most ideal for mine drainage treatment processes. This assembly allows free access and easy daily observation of electrode condition.

Electrode fouling has long been a problem with mine drainage pH control systems. Self-cleaning accessories for electrodes, however, are available from many vendors. These cleaning mechanisms can be a membrane or nylon brush or an ultrasonic transducer.

The membrane wiper is employed where electrode filming occurs, perhaps from ferrous or ferric iron. The membrane has a reciprocating action (up-and-down motion) across the measuring electrode. This will remove buildup of material that can cause an error in measurement.

Another electrode cleaner is a nylon brush, which functions basically like the membrane wiper, except that it is used where crusting solids (calcium carbonate or calcium sulfate) deposit on the electrode.

These devices should be considered an auxiliary method for a more reliable pH control system and should reduce the amount of maintenance required for clean probes. Nevertheless, pH electrode cleaning remains a daily maintenance duty. Therefore, provisions should be made by the designer for easy access to the probes and with certain consideration for operation personnel. Probes should not be positioned in confined areas or places that will expose personnel to unnecessary dangers.

The immersion pH assembly is found more often in mine drainage plants than any other type, and is positioned in an open transfer trough where the measured liquid maintains scour velocity. If this type of arrangement is impossible, a small auxiliary control loop in conjunction with a pump can be used. More simply, a recirculation loop with flow-through pH electrodes has produced a good process loop with reliable performance.

The placement of pH electrodes inside the neutralization or flash mix tank should be avoided. If unavoidable, certain provisions such as a still well for the probes should be provided. An example of an arrangement is shown in Figure 8-1.

When a pH probe is used as the primary control device, it is highly recommended that it be positioned between the flash mix tank and the subsequent treatment processes. Excellent control was observed by Wilmoth by placing the pH probe in the inlet of the clarifier, thus minimizing problems of probe fouling.

Many engineers/designers believe that a variable-speed lime feeder integrated with a pH controller can be used to pace the addition of neutralizing agent. Although this has been done successfully, it is subject to failure in highly variable situations (extreme fluctuations in acidity), where the pH controller has insufficient reaction time to changes in acidity. This "lag time" causes the feeder to fall behind--sometimes to a point where it is impossible to correct itself.

More successful systems have utilized on-off lime feeder arrangements with enlarged flash mixing tank to provide a better buffering zone. Many existing mine drainage plants utilize 1- to 3-minute detention volumes. However, larger mixing tanks with 30-minute to 1-hour detention volumes have been

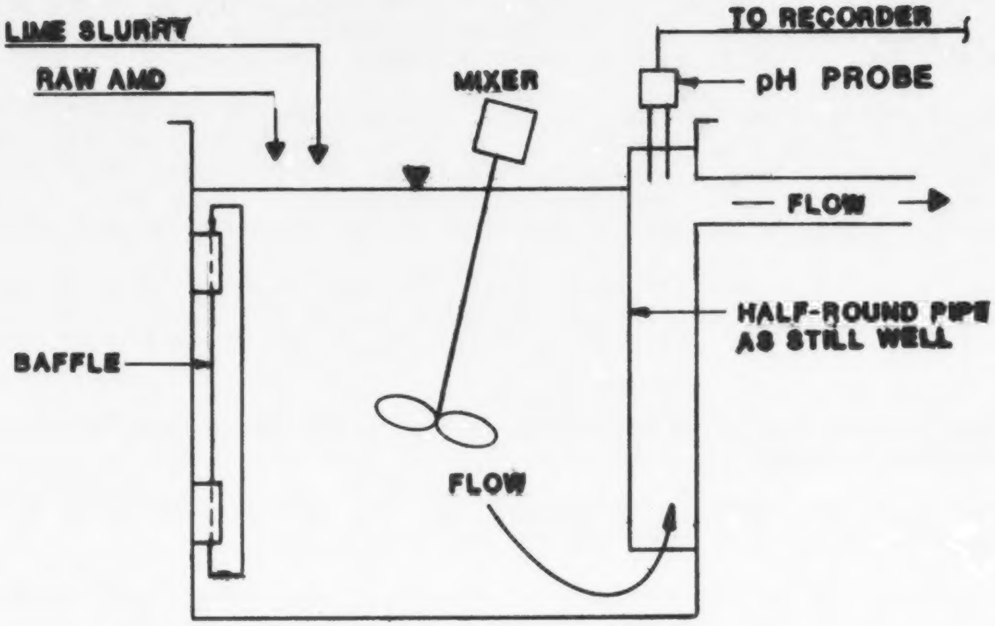

Figure 8-1. Flash mix tank with pH probe installed.

employed. This reactor allows more pH control reaction time and a steadier state operation.

8.5 Level Controls

Level controls are incorporated into many of the processes used in mine drainage treatment. Mercury float controls are used with the raw water pumps, solid level indicators in the silo, and sensing probes or sonic level controls in the slurry or stabilization tank.

Most raw water pumps, generally in the mine, are activated by hand and run constantly. Sometimes, the equalization basin has transfer pumps that utilize mercury controls or the conventional float with rod.

When using a lime silo, at least a high-level and a low-level indicator should be installed. The exact position of these probes is up to the designer's discretion. The low-level indicator, however, should allow at least 1 or 2 days supply of lime when activated. This precaution will allow continuous operation between deliveries.

Perhaps the best noncontact silo level indicator is the strain gauge. These microcells are placed on the legs or skirt of the silo and measure the deflection of the steel. Equipped with a recorder, a display in the percent full gauge can be read with remarkable accuracy. These strain gauges offer an excellent choice as a solids level indicator.

The slurry or stabilization tank will probably contain the principal level controls. These high and low levels, along with alarms, usually control other equipment such as slurry pumps and lime feeder. These level controls perhaps represent the most difficult application. In the past, a set of physical probes (rods) have performed well, but required frequent cleaning.

Today's market offers the sonic level indicator that can be programmed to control system equipment. These noncontact devices are an excellent alternative for nonconfined (i.e., trough, large mixing tank) liquid level measurement.

These devices have performed poorly in lime silos, where "echoing" produces erroneous readings.

Many types of controls can be employed. The decision becomes a designer preference based upon experience and past success.

In summary, an initial awareness of electrical requirements with respect to voltage, phase, transformers, and other hardware should alleviate later complications and save costly construction time. Also, the instrumentation and process control panel should be isolated from the active operation; ideally, it should be placed in a separate building. The small additional capital investment for this system will have a hundredfold return through a cleaner, healthier environment for maintenance personnel, which will result in better operation and longevity of equipment.

CHAPTER 9

REVERSE OSMOSIS

9.1 Introduction

The application of reverse osmosis to the treatment of AMD has been extensively studied by the U.S. Environmental Protection Agency over the last decade. These studies have demonstrated that reverse osmosis (RO) can be highly effective in removing most of the dissolved solids in acid mine drainage. The current purpose of applying reverse osmosis to AMD treatment is to produce a potable water while achieving maximum recovery. The product water will be low in dissolved solids, usually less than 100 mg/l, but may contain chemical or bacterial constituents that exceed drinking water standards. Reverse osmosis product water is not initially acceptable as potable water, but it is of excellent quality and should be considered for other uses, including boiler feedwater, cooling water, bathhouse shower water, or for a variety of other industrial purposes. Although feedwater recoveries of 60% or more can be obtained, design recoveries of 50% are more practical since the emphasis is on producing specific quantities of high-quality effluents and not on treating the entire volume of AMD by reverse osmosis.

This chapter is intended to provide the basic design criteria and operating parameters applicable for using the reverse osmosis process. Using these criteria and operating parameters, the designer can evaluate this process for a particular application. It must be emphasized that reverse osmosis is a complicated process as compared to other methods of treatment; therefore, pilot testing must be performed to establish the design parameters as well as pre- or post-treatment needs.

9.2 Operational Considerations

Osmosis occurs if two solutions of different concentrations in the same solvent are separated from one another by a membrane. If the membrane is semipermeable (i.e., permeable to the solvent and not to the solute), then the solvent will flow from the more dilute solution to the more concentrated solution until an equal concentration results. In reverse osmosis, the direction of solvent flow is reversed by the application of pressure to the more concentrated solution. As a result, the concentrated solution, termed the solute or brine, becomes more concentrated. The solvent, termed the permeate, is the product from the process.

9.2.1 Membrane Type and Configuration

Tubular, hollow-fiber, and spiral-wound membrane types have been tested for use in treating acid mine drainage. Studies performed by Wilmoth (1) in 1972 indicated that the spiral-wound configuration with a formamid-modified cellulose acetate membrane was slightly superior to others with respect to the average flux (permeate flow rate), long-term flux stability, and dissolved solids rejection. Since then new membrane materials have become commercially available. It is recommended that these be treated in the spiral-wound configuration for any RO application.

9.2.2 Pretreatment

Problems with membrane fouling can occur as the concentrations of various compounds increase during the process. Most important is the potential for iron foulng and calcium sulfate (gypsum) formation. Iron fouling has been minimized by lowering the feedwater pH, or flushing the RO membrane by operating at lower pressures for short periods. When the raw AMD contains high concentrations of sulfates, gypsum can form if its solubility is exceeded. In this case, this process may not be applicable.

9.2.3 Prefiltration

Cartridge or bag filters should be used to minimize fouling by suspended solids in the feedwater. This can increase membrane life and improve rejection levels. The filters should be capable of removing particles larger than 20 μm. The filters are placed at the suction side of the RO feed pumps. Duplicate units should be provided in parallel to eliminate the need to shut down in the RO system when cleaning or filter replacement is necessary.

9.2.4 pH Control

In treating AMD, the pH of the feed should be maintained between 2.8 and 3.0. Adjustment of the feed pH is required to prevent the precipitation of slightly soluble inorganic salts such as calcium and iron. At a pH less than 3.0, ferric iron (Fe^{3+}) remains dissolved. When pH values exceed 3.0, ferric hydroxide may begin to precipitate on the membrane surface.

Although a low pH is necessary to improve operating conditions when treating AMD, it is lower than the optimum range of 5.0-6.5 for the cellulose acetate membranes. The life of this common membrane formation will be decreased, but newer, pH-resistant membranes are now available.

9.2.5 Disinfection

Another method used to reduce iron fouling problems is to provide disinfection to inhibit microbial activity in the raw mine drainage feed. Ultraviolet light, proven to be an effective bactericide, is recommended to prevent an accumulation of iron-oxidizing bacteria on the membrane surface.

9.2.6 Space Requirements

In the design of any system, sufficient space must be available to accommodate the entire system, including all major mechanical and electrical equipment, auxiliary equipment, and storage facilities. The initial design should consider the ease of installation and modularity, or the ability to add modules, stages, or racks to the system as needed.

9.2.7 Design Capacity

A loss in flux or permeate flow rate, due to compaction or chemical fouling of the membrane surface, will gradually reduce the production capacity of the system. This should be offset by appropriate sizing during the initial design. The RO system should be sized to process the daily AMD flow in 20 hours at the average flux rate. This provides an adequate safety margin for processing the daily flow and sufficient time for daily maintenance and membrane cleaning. The designer should also consider providing storage capacity in the event the system is out of service for a prolonged period. This can be accomplished during the initial design phase by oversizing feedwater holding tanks or by including larger storage ponds. There is no specific storage capacity design value, but a volume of several days flow should be considered.

9.2.8 Flux Rate--Feed, Product, and Concentrate

One of the design factors critical to a successful RO operation is an accurate permeation rate (flux rate) over the life of the membranes. This is essential to estimate the quantity of installed surface area, cleaning cyles, and membrane replacement. Initially, with the feed rate constant, a decline of permeate flow will occur due to membrane fouling. Even without fouling, a slight flux decline will be observed because of membrane compaction.

The system should be designed to produce a constant permeate output based on the daily design flow. This is normally accomplished by using pressure-compensating flow controls that automatically adjust for flow variations. Once preset, the control valve automatically adjusts the operating pressure to maintain the permeation rate at its predesignated flow rate.

Tests were performed on a 15.14-m^3/d (4,000 gal/d), once-through, continuous operation; an average flux rate of 500 $l/m^2/d$ (12.3 $gal/d/ft^2$) operating at 28.2 atm (400 lb/in^2g) and 75% recovery was realized.

It is difficult to state a specific flux rate for design purposes, since flux varies with operating pressure, concentration of the feed stream, and overall recovery. Perhaps one of the best methods for determining a flux rate is to extrapolate annual flux values from pilot tests and apply these to the design of a full-scale system.

9.2.9 Operating Pressure

The system should be designed to operate at 281 kg/cm^2 (400 lb/in^2). At this operating pressure, minimum membrane compaction will be experienced while maintaining adequate flux ratios to assure high effluent quality. (Effluent quality will decrease as operating pressure decreases.)

As the flux declines, the pressure control system will compensate for the decreased flow by increasing the operating pressure, thus maintaining a constant product flow. A loss in flux, due to compaction, is typically offset by appropriate sizing and startup at a reduced pressure with gradual increase in operating pressure over the life of the modules. To minimize fouling, each module should operate with a 10:1 brine-to-produce flow ratio. This brine velocity presents "boundary layer" development, which is a layer of stagnant water against the membrane surface.

9.2.10 Module Configuration

The type of pressure vessel manifolding arrangement (series or parallel) is dictated by the desired recovery level and the need to maintain an adequate brine-to-product flow ratio. Pressure vessel arrangements and modules are designed so the raw feed enters a parallel bank of pressure vessels, and the concentrate from this bank is used as the feed for the next parallel arrangement of vessels.

In high-recovery continuous flow systems, it is advantageous to design the system so only a small number of modules have to process the most concentrated portion of feed stream. Then if fouling due to chemical precipitation occurs, it is confined to a minimum number of modules.

In research and testing done at the EPA Crown Mine Drainage Control Field Site, the vessel array configurations and corresponding recovery levels were as follows (1):

Arrangement	Recovery Level %
8-6	40-60
7-4-3	60-75
5-4-3-2	75-80
5-4-2-2-1	85-90

At high recovery levels of 85%-90%, only one pressure vessel (6 modules) would be subjected to the most concentrated material. Should precipitation occur, only these modules would be severely fouled.

9.2.11 Safe Recovery Levels

There are no established design parameters for determination of ultimate recovery levels. Rather, two controlling factors limit the overall recovery of water from the treatment process.

The first is the precipitation of calcium sulfate ($CaSO_4$). Acid mine waters typically contain high concentrations of calcium and sulfate ions. As the RO process progresses, calcium and sulfate concentrations increase. Once the solubility of calcium sulfate is exceeded, precipitation occurs. Wilmoth determined the empirical limit for RO recovery levels with AMD to be $R = 100 - 0.55 (Ca \times SO_4)^{-\frac{1}{2}}$ where Ca and SO_4 concentrations (mg/l) are those used in the raw feed. It is well known that gypsum (calcium sulfate) is only slightly soluble. When concentrated, it will precipitate and form a hard, tenacious scale on tanks, piping, and, more importantly, the membranes.

The second factor influencing recovery rates is the desired quality of the permeate. As the drainage is processed, the concentration of total dissolved solids (TDS) in the permeate increases almost linearly with the TDS in the concentrate. Thus, increasing recovery increases the concentration of pollutants in the waste (reject) stream and in the product water. The final use of the permeate determines the maximum recovery of the process.

9.2.12 Dissolved Solids Rejection

The percentage rejection of waste stream contamination is greater than 90%; in most cases, the spiral-wound cellulose acetate membranes will reject 99% or more of the dissolved salts in the raw AMD feed. Table 9-1 shows anticipated permeate water quality.

As the pollutants in the reject are concentrated, more dissolved solids contact the membrane; thus, more will pass through the membrane and deteriorate the permeate quality.

9.2.13 Membrane Life Expectancy

Membrane life is affected by pH, temperature, and operating pressure. One manufacturer defines useful membrane life as the time taken for the membrane to lose 40% of its initial flux. Any attempt to accurately determine how long a membrane will last under normal operating conditions must be accomplished through long-term studies. Under applications of this nature, manufacturers project a cellulose membrane life of 3 years. For operating cost projections, it is suggested that the designer assume that 50% of the membrane area will be replaced each year.

9.2.14 Concentrate Treatment and Disposal

As the RO system removes dissolved solids, the process generates a highly concentrated waste stream that requires treatment before disposal. The exact

TABLE 9-1

ANTICIPATED PERMEATE WATER QUALITY

Parameter[a]	Raw Water Quality	Product Water Quality
pH (units)	3.4	4.3
Specific conductance (mhos)	1,020	32
Acidity	210	32
Calcium	150	1.2
Magnesium	115	1.4
Iron, total	110	1.2
Iron, ferrous	71	0.8
Aluminum	15	0.8
Manganese	43	0.4

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[a] All values expressed as mg/l unless otherwise noted.

volume and salt content of the concentrate stream depends on the influent quality as well as the recovery rate. For example, an RO system with a 90% recovery rate creates a waste stream with a pollutant concentration 10 times that of the feedwater, but with a volume of only 10% of feed. This concentrate must be treated and/or disposed of in some environmentally acceptable manner. The possible treatment and disposal methods include the following:

1. lime neutralization;
2. evaporation - mechanical and/or atmospheric;
3. contract disposal.

Lime neutralization of the waste stream from the RO process is a practical disposal and treatment method. This method is adequately described in this manual.

Possible evaporation techniques include the mechanical, wiped-film unit, which is capable of reducing the volume by 75% or more. In drier parts of this country, evaporation ponds can be considered. These provide for atmospheric evaporation to reduce the volume, possibly to dryness.

Contract hauling and disposal by an approved waste hauling firm is another alternative.

9.2.15 Capital and Operating Costs

Accurate cost data relative to RO treatment of acid mine drainage are virtually nonexistent. Extensive studies of actual commercial units necessary to develop these costs have never been performed. Capital and operating costs presented in the literature are only estimates based on system size and degree of sophistication. Operating costs are also estimates, for the most part based upon assumed values.

9.3 References

1. Wilmoth, R.C. Applications of Reverse Osmosis to Acid Mine Drainage Treatment. EPA-670/2-73-100, Environmental Protection Technology Series, Cincinnati, Ohio, December 1973.

9.4 Other Selected Readings

Rex Chainbelt, Inc. Reverse Osmosis Demineralization of Acid Mine Drainage. Program No. 14010 FQR, U.S. Environmental Protection Agency, Water Pollution Control Research Series, March 1972.

Gulf Environmental Systems Company. Acid Mine Waste Treatment Using Reverse Osmosis. Program No. 14010 DYG, U.S. Environmental Protection Agency, Water Pollution Control Research Series, August 1971.

CHAPTER 10

ION EXCHANGE

10.1 Introduction

Ion exchange, like reverse osmosis, can be utilized to treat acid mine drainage for the removal of unwanted dissolved ions to produce a water of excellent quality for many industrial uses. Ion exchange can also produce a potable grade of water; but such a system would have to be followed by filtration and disinfection to comply with public health regulations.

Ion exchange in water treatment is defined as the reversible interchange of ions between a solid medium and the aqueous solution (1). To be effective, the solid ion-exchange medium must contain ions of its own, be insoluble in water, and have a porous structure for the free passage of the water molecules. Within the solution and the ion-exchange medium, a charge balance or electroneutrality must be maintained; i.e., the number of charges, not the number of ions, must stay constant. Ion exchange materials usually have a preference for multivalent ions; therefore, they tend to exchange their monovalent ions. This reaction can be reversed by increasing the concentration of monovalent ions. Thus, a means exists to regenerate the ion exchange material once its capacity to exchange ions has been depleted (2, 3).

The most common ion exchange use is the softening of "hard" or mineral-bearing water for domestic or commercial purposes. The hardness in water is attributed to its calcium and magnesium content. Initially, the ion exchange material is charged with monovalent cations, usually sodium (sodium chloride). The hard water is passed through a bed of ion exchange material, and the divalent calcium and magnesium cations are exchanged for sodium ions as follows:

$$Ca^{++} + 2Na^{+}\ (resin) \rightarrow Ca^{++}\ (resin) + 2Na^{+} \qquad (30)$$

Ion exchange materials tend to form stable compounds through this exchange principle. When more than one type of cation is available, the material will have an affinity for certain ones more than others. In commercial or industrial applications, the ion exchange resin is usually operated in the proton (H^{+}) or acid cycle. Here, sodium is replaced with a proton (H^{+}) and the exhausted resin is regenerated with sulfuric or hydrochloric acid.

The earliest ion exchange materials were either natural or synthetic zeolites--a mineral produced from mixtures of aluminum salts and silicates. In the 1930's, plastic materials called resins were developed that expanded the applications of ion (cation) exchange. An anion exchange resin was developed in 1949 that enabled the process to be used for total demineralization of water. In the present-day technology of ion exchange, the resins available can be classified as strong-acid cation, weak-acid cation, strong-base anion, and weak-base anion types. Combinations of the available resins have been used in systems for treatment of different waters for specific purposes (4, 5).

The application of these ion exchange systems for the treatment of mine drainage has been studied mainly to produce potable water where a reduction in the total dissolved solids is required. Processes developed include the Sul-biSul Process, the Modified Desal Process, and the Two Resin Process. The operation and performance of the first two of these processes have been demonstrated in full-size facilities; the latter process, in pilot units. It has been concluded that ion exchange can be used to demineralize mine drainage and produce water with a quality acceptable for potable or industrial use, but the costs of operation do not appear competitive with other methods.

10.2 Sul-biSul Process

The Sul-biSul Process was developed by the Nalco Chemical Company but is now assigned to the Dow Chemical Company (6). The process employs a two- or three-bed system, depending upon the mine drainage quality. Cations are removed by a strong-acid resin in the hydrogen form, or by a combination of weak-acid and strong-acid resins (7, 8). The AMD feed is first passed through the cation exchanger, which removes the metal cations and exchanges these for hydrogen protons, or (H^+) ions. This reaction is expressed by Equation 31.

$$Fe^{+2}SO_4 + 2HR \rightarrow MR_2 + H_2SO_4 \qquad (31)$$

where R represents the strong-acid exchange groups on the resin, and M represents a divalent metal cation, such as iron (ferric), calcium, or manganese.

The product water from this first exchange contains additional sulfuric acid from the displaced proton (H^+). Following this, the water is decarbonated to remove carbon dioxide formed during the cation exchange process. Then a strong-base anion resin (R') operating in the sulfate-to-bisulfate cycle removes both the sulfate and hydrogen ions during this exchange reaction (Equation 32):

$$R'SO_4 + H_2SO_4 \rightarrow R'(HSO_4)_2 \qquad (32)$$

Because of the high acidity of the feed, sulfate ions in solution and on the resin are converted to the bisulfate form. This conversion of bivalent sulfate to monovalent bisulfate provides for twice the amount of sulfate to be stored on the resin. Removal of the sulfate results in good-quality water. Regeneration of the cation exchange bed is accomplished with either hydrochloric or sulfuric acid. In the regeneration of the anion bed, bisulfate ions are converted back to the sulfate form by the feedwater in a reversal of Equation 31. The addition of lime slurry to the regenerant will speed this reaction. The unusual feature of this process is the removal of sulfate from feedwater by anion exchange using only water or water with a little added alkali as the regenerant (9). The product water must be filtered and chlorinated according to public health regulations before use as a potable water. Wastes from the regeneration process would have to be treated before discharge.

The Sul-biSul Process can be used to demineralize brackish water containing predominantly sulfate anions with a dissolved solids content up to 3,000 mg/l. The raw water should have an alkalinity content about 10% of the total anion concentration and a sulfate-to-chloride ion ratio of at least 10:1. The process is especially suited to alkaline waters containing calcium sulfate, such as those contaminated by mine drainage (9).

Limitations of the process center on the low exchange capacity of the anion exchange resin and its inefficient method of regeneration. The exhausted anion resin can be regenerated by the raw water itself; however, this requires a considerable volume of water and takes a significant length of time if the sulfate content is low. The addition of a cheap alkali such as lime is reported to improve the regeneration; however, one study showed poor results (10).

One problem is the requirement for disposal of this large volume of regenerants. This water must be sufficiently alkaline and abundant so that it can be used as the regenerant and then discharged to the stream. If the raw water cannot be used as the anion bed regenerant, other alkalis must be employed. When this is necessary, tests have indicated that there may be a negative net production of water and the process may not be economically competitive (9).

A water treatment plant using this process began operation in 1971 at Smith Township, Pa. The plant was designed for the production of 1,900 m^3/d (0.5 Mgal/d) of potable water. The raw water supply at Smith Township is a small stream that is affected by mine drainage. Studies indicated that it contained sulfates, iron, and manganese, yet remained alkaline. Pilot tests confirmed the applicability of the Sul-biSul Process (11). Projected raw and finished water quality for this plant are shown in Table 10-1, and a flow schematic of the system in Figure 10-1 (12).

Cost data for the Sul-biSul Process are limited to the projections from a few studies and the Smith Township plant. At Smith Township, a continuous ion exchange-regeneration system was installed at a capital cost of $898,000. Operating costs are not available because the plant did not meet design capacity specifications and is not being operated pending litigation.

Projected operating costs were estimated to be in the range of \$0.10-\$0.13/m^3 ($0.40-$0.50/1,000 gal). These costs have more than doubled if extrapolated

TABLE 10-1

PROJECTED RAW AND FINISHED WATER QUALITY
SUL-BISUL PROCESS AT SMITH TOWNSHIP, PA.

Parameter	Typical Quality	
	Raw Water	Finished Water
pH	6.5 - 8.4	8.0
Alkalinity, mg/l	76	10 - 30
Dissolved solids, mg/l	1,500 - 2,000	300
Sulfates, mg/l	400 - 1,300	50 - 100
Hardness, mg/l	1,600	150
Chlorides, mg/l	16	2

to a 1980 cost basis. Unless a sufficient supply of excessively alkaline water is available for regeneration of the anion resin, the Sul-biSul Process cannot economically produce potable water from acid mine drainage (9).

10.3 Modified Desal Process

The Modified Desal Process is another ion exchange process that has been investigated for treatment of AMD to recover potable water (8, 9, 10, 13). This process uses a weak-base anion resin in the free-base form, which is converted to the bicarbonate form to treat the raw AMD. The weak-base resin exchanges sulfates (or other anions) for bicarbonate, allowing the cations to pass through the bed according to the following reaction:

$$MSO_4 + 2R''HCO_3 \rightarrow R''_2SO_4 + M(HCO_3)_2 \qquad (33)$$

where R'' = is the weak-base exchange group on the resin matrix

M = a divalent metal ion

The solution of metal bicarbonates is aerated to oxidize ferrous iron to ferric iron and to purge the carbon dioxide gas. The effluent is then treated with lime to precipitate metal hydroxides, settled to remove suspended solids, and then filtered and chlorinated if it is to be used as a potable water (7, 9, 13).

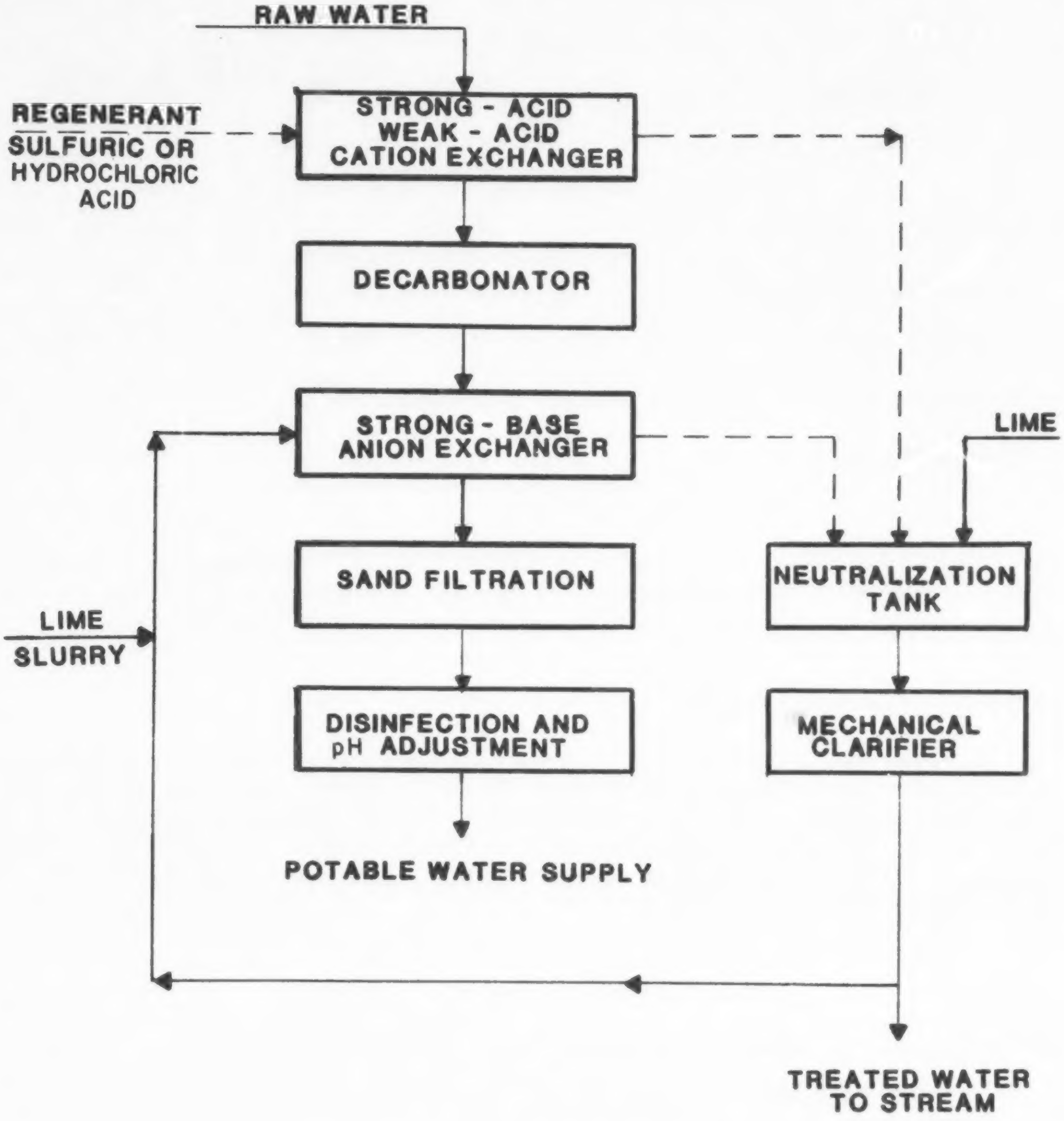
RAW WATER
REGENERANT
SULFURIC OR
HYDROCHLORIC
ACID
STRONG - ACID
WEAK - ACID
CATION EXCHANGER
DECARBONATOR
STRONG - BASE
ANION EXCHANGER
LIME
SAND FILTRATION
NEUTRALIZATION
TANK
LIME
SLURRY
DISINFECTION AND
pH ADJUSTMENT
MECHANICAL
CLARIFIER
POTABLE WATER SUPPLY
TREATED WATER
TO STREAM

Figure 10-1. Sul-biSul Process continuous ion exchange flow sheet.

Ammonia is used as the alkaline regenerant to displace sulfate from the exhausted resin. Lime is used to precipitate the ammonia regenerant for reuse. In this way, ammonia is recycled in the process. It is possible to recover the carbon dioxide and lime used in this process by roasting lime sludge wastes in a kiln. If this were done, all of the principal chemicals used in the process would be recycled. The net result would approach a zero discharge process, with fuel, power, and makeup quantities of chemicals fed into the plant, and only potable water, iron hydroxide, and calcium sulfate sludge discharged from it (9).

The Modified Desal Process is not limited by total dissolved solids or pH levels; however, large quantities of carbon dioxide are required to achieve good resin utilization for high total dissolved solids or alkaline feed waters. The process is limited in application to waters containing less than 2,200 mg/l of sulfate. Another limitation is that mine waters containing iron in the ferric form may cause fouling of the anion bed because of precipitation of ferric hydroxide (9, 13).

A demonstration plant for treatment of AMD by the Modified Desal Process was constructed in 1972-73 by the Pennsylvania Department of Environmental Resources at Hawk Run near Philipsburg, Pa., at a capital cost of $2,335,000. The purpose of this plant was to demonstrate the applicability of the Modified Desal Ion Exchange Process for treating acid mine drainage. Its secondary purpose was to provide a drinking water supply with a capacity of 1,892.5 m^3/d (0.5 Mgal/d) for the nearby Philipsburg area. A typical water analysis is shown in Table 10-2.

Operating costs in 1975 were $118,925 or $0.14/$m^3$ ($0.54/1,000 gal) of plant capacity based on an output of 2,271 m^3/d (0.6 Mgal/d). This operating cost does not include depreciation of the capital cost, which was estimated to be $214,000 annually. It should be pointed out that this plant was designed and operated as a research facility, so all costs are higher than would be expected in a normal production facility (13). Estimated operating costs for a 750 mg/l sulfate feed were $0.49/$m^3$ ($1,85/1,000 gal) including depreciation (9). The 1975 actual costs of water produced by the Modified Desal Process were about $0.48/$m^3$ ($1.82/1,000 gal) when depreciation was included.

Several optimizing modifications have been made at the Hawk Run Plant to increase its efficiency. An important one involves precarbonating the acid mine drainage, which has resulted in a significant increase in capacity from 1,893 to 3,028 m^3/d (0.5 to 0.8 Mgal/d). A schematic of this process at Hawk Run is shown in Figure 10-2.

The waste regenerant is composed of an ammonium sulfate solution. This is lime-treated to form calcium sulfate, which is then removed by filtration. The filter effluent is sent to a distillation process where 92%-95% of the ammonia is recovered for reuse as the first-stage regenerant.

The Hawk Run Plant was constructed to demonstrate the process for augmentation of a degrading water supply. Water quality has now improved to the point that the plant is currently not needed, and it has been placed in the standby mode pending its future use. The plant has also been offered for

TABLE 10-2

TYPICAL WATER ANALYSIS
HAWK RUN AMD TREATMENT PLANT

	Raw Water	Product Water
Hot free mineral acidity (mg/l $CaCO_3$)	384	--
Free mineral acidity (mg/l $CaCO_3$)	362	--
Iron (mg/l)	101	0.2
Calcium (mg/l $CaCO_3$)	295	85
Magnesium (mg/l $CaCO_3$)	100	99
Total hardness (mg/l $CaCO_3$)	395	184
Sulfate (mg/l)	684	192
Total dissolved solids (mg/l)	1,084	284
pH (Standard Units)	3.7	9.5

sale to the nearby community for $1, but it was not accepted because of its high operating cost compared to other water treatment processes (13). While it was operated, it performed extremely well (14).

10.4 Two Resin Process

In 1972, the Culligan International Company (10) investigated a standard two-resin system. This process has been further investigated by the U.S. Environmental Protection Agency at their Crown, W. Va., Research Facility (14).

The Two Resin Process involves the use of a strong-acid cation exchanger in the acid (H^+) form followed by a weak-base anion exchanger in the free-base (OH^-) form as shown in Figure 10-3. In the cation column, protons (H^+ ions) are exchanged for the metal ions in the acid mine drainage. Following cation exchange, the anion column feed is rich in sulfuric acid.

Total metal cation removal greatly increases the regenerant dosage and the operating cost of the system. It has been demonstrated that significant cost

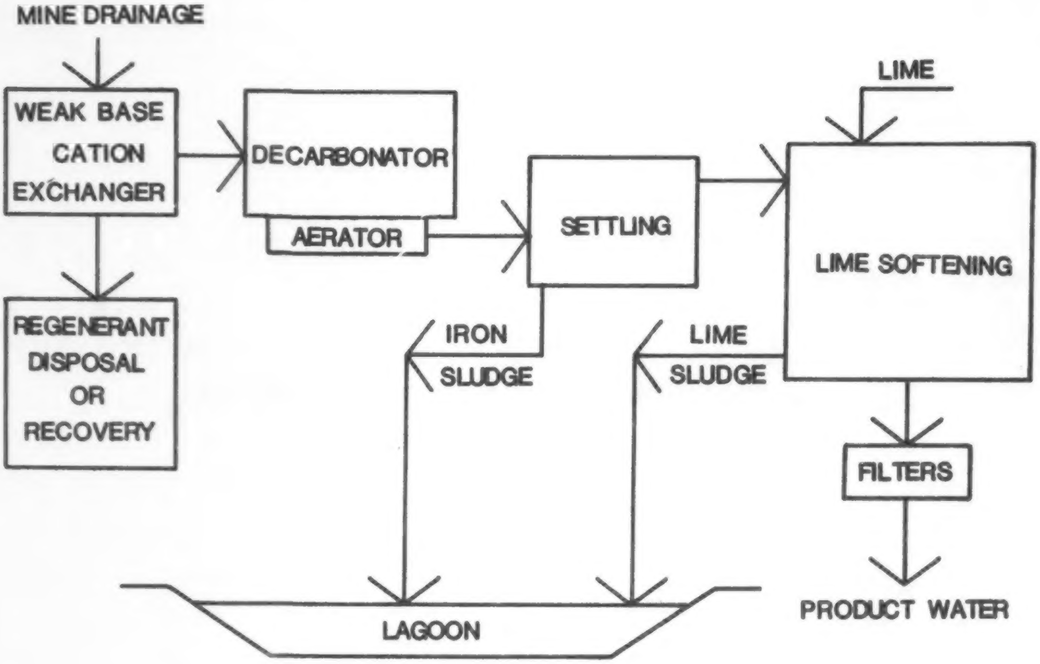
MINE DRAINAGE
WEAK BASE CATION EXCHANGER
DECARBONATOR
AERATOR
SETTLING
LIME
LIME SOFTENING
REGENERANT DISPOSAL OR RECOVERY
IRON SLUDGE
LIME SLUDGE
FILTERS
LAGOON
PRODUCT WATER

Figure 10-2. Modified Desal Process flow diagram.

reductions can be realized by operating the system consistent with the needs for product water end-use. The concentration of residual metals in the cation exchanger effluent can be optimized by controlling the dosage of regenerant.

Feed to the anion exchanger is predominantly sulfuric acid, which is totally absorbed by the resin. A weak-base anion exchange resin only absorbs acids; it cannot split neutral salts. The anion exchange effluent is alkaline, and some precipitation of residual iron and aluminum ions can be expected. The effect of this accumulation on the anion resin efficiency and capacity must be monitored, but was observed by Wilmoth to have no significant adverse effects in 900 regeneration cycles.

Either sulfuric acid or hydrochloric acid may be used for regenerating the cation exchanger. Sulfuric acid is usually preferred because of its lower cost; however, gypsum may form if sulfuric acid is used. If so, H_2SO_4 regenerate concentration must be limited to 2% by weight. Treatment for disposal of both regenerant streams is necessary.

Water quality results from the EPA Crown studies are summarized in Table 10-3. As could be expected. an increase in the cation regenerant dosage

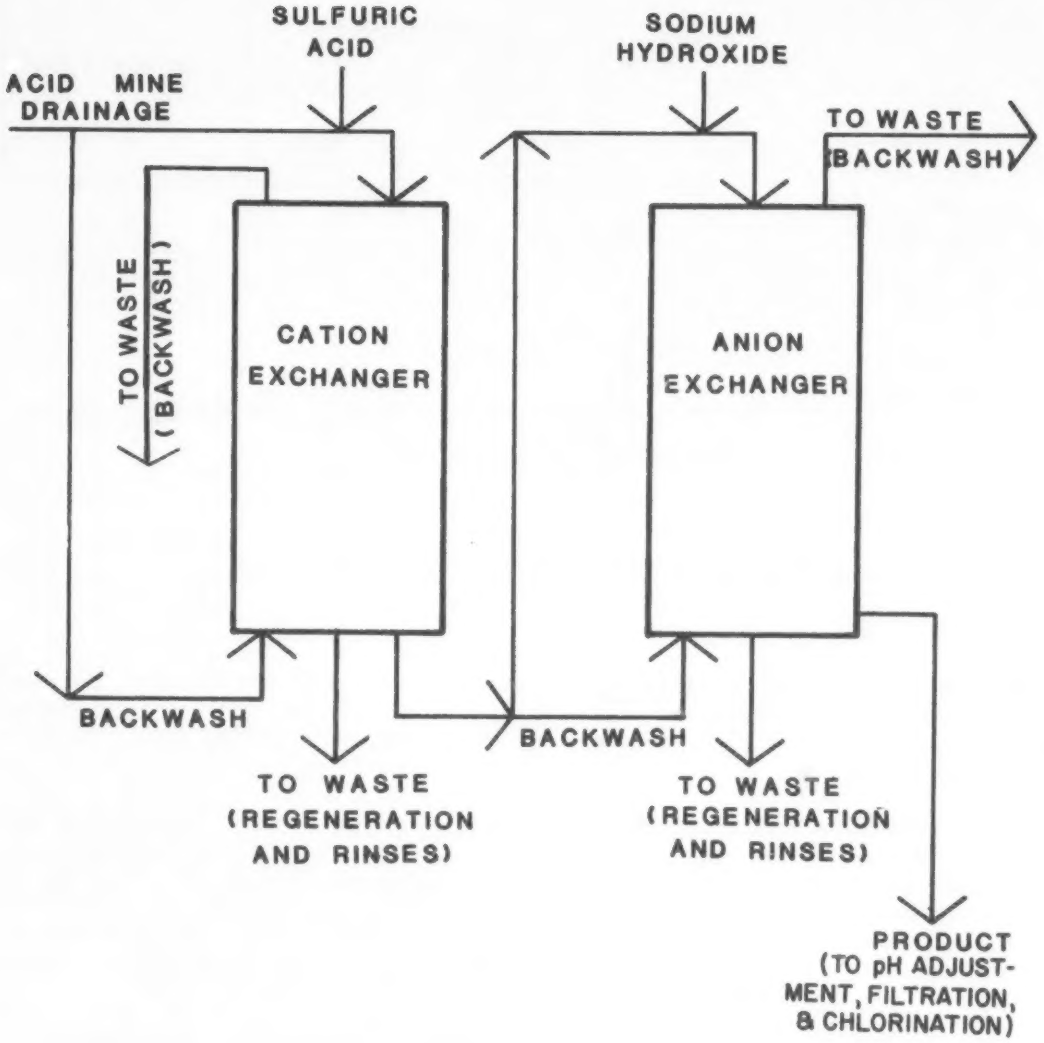
SULFURIC ACID
ACID MINE DRAINAGE
SODIUM HYDROXIDE
TO WASTE (BACKWASH)
TO WASTE (BACKWASH)
CATION EXCHANGER
ANION EXCHANGER
BACKWASH
BACKWASH
TO WASTE (REGENERATION AND RINSES)
TO WASTE (REGENERATION AND RINSES)
PRODUCT (TO pH ADJUSTMENT, FILTRATION, & CHLORINATION)

Figure 10-3. Two-resin ion exchange system.

TABLE 10-3

SUMMARY OF ION EXCHANGE SYSTEM CHEMICAL ANALYSES[a]

Sample	Cond	Acid alk[b]	pH	Ca	Total Iron	Fe^{2+}	Na	Al	Mn	SO_4	TDS
OPERATION AT 48-g/l (3-lb/ft^3) DOSAGE OF SULFURIC ACID											
Raw feed	2,870	410	4.9	400	210	190	360	9.3	5.0	2,600	2,690
Cation effluent	8,890	1,780[b]	1.54	42	23	22	260	0.6	0.6	2,460	2,790
Anion effluent	1,230	280	9.4	38	3.1	0.3	320	0.1	0.4	610	980
OPERATION AT 96-g/l (6-lb/ft^3) DOSAGE OF SULFURIC ACID											
Raw feed	2,870	410	5.1	330	200	190	340	8.7	5.0	2,430	3,410
Cation effluent	8,910	1,730[b]	1.55	29	14	13	240	0.6	0.3	2,380	2,670
Anion effluent	1,370	340	9.3	21	2.4	0	380	0.3	0.2	570	970
OPERATION AT 144-g/l (9-lb/ft^3) DOSAGE OF SULFURIC ACID											
Raw feed	2,790	430	5.0	340	210	200	350	7.8	5.2	2,470	3,500
Cation effluent	9,700	2,000[b]	1.58	24	16	13	190	0.7	0.8	2,280	2,550
Anion effluent	1,430	290	9.5	19	1.4	0	400	0.5	0.2	730	1,150
LONG-TERM OPERATION AT 48-g/l (3-lb/ft^3) DOSAGE OF SULFURIC ACID											
Raw feed	2,770	500	4.5	350	180	170	330	8.5	5.3	2,440	3,420
Cation effluent	8,220	1,930[b]	1.61	45	22	19	250	0.74	0.62	2,370	2,700
Anion effluent	1,310	290	9.3	38	6.7	1.1	330	0.21	0.44	660	1,050

---

[a] All units are expressed as mg/l except for conductivity (μmhos/cm) and pH (standard units).
[b] Acidity, alkalinity, and exchangeable cations are expressed as $CaCO_3$.

resulted in a decrease of the divalent cations in the exchanger effluent. The raw AMD was found to be very high in sodium, but very little removal of this monovalent ion was achieved. The anion exchange column effectively removed all acidity and imparted alkalinity. Precipitation of iron within the column was also observed, but no deleterious effects could be documented.

Long-term testing was performed at minimal cation regenerant dosage. A deterioration in the performance of the cation column was observed. Although the utilization efficiency of the regenerant decreased, the product water quality remained acceptable. Unexpectedly, there was no apparent reduction in the anion exchanger efficiency, even though iron precipitation did occur. Using the low-pH cation column effluent for backwashing the anion column effectively prevented the iron hydroxide precipitate from inhibiting anion column performance.

Except for the presence of sodium ions, the water produced at Crown would meet potable requirements following filtration and chlorination. Estimated capital costs for plants utilizing the Two Resin Process on a 1978 cost basis range from $1,000,000 for a 1,893-$m^3$/d (0.5 Mgal/d) plant to $1,700,000 for a facility capable of producing 3,785 m^3/d (1.0 Mgal/d) (10). For the Crown raw water, which is more polluted than most AMD, operating costs appear to be about 50% greater than those for the Modified Desal Process.

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CHAPTER 11

CHEMICAL SOFTENING

11.1 Introduction

Chemical softening is employed as a treatment process to remove dissolved ions from AMD only when considering the effluent for industrial use or possibly as potable water. Softening processes have been used extensively to remove hardness (calcium, magnesium, iron, manganese, and aluminum) from municipal water supplies, and can be adapted to treat acid mine drainage.

Two processes that merit consideration for possible potable water production are lime-soda and alumina-lime-soda. The former has been the most common process used by water plants to treat hard water. The latter was developed during 1971 as a brackish water desalination process (1). The advantages chemical softening has over other water recovery methods--e.g., ion exchange or reverse osmosis--are that only conventional treatment equipment is required and no waste streams other than sludge are produced.

11.2 Lime-Soda Softening Process

The lime-soda softening process has been widely used to remove hardness, iron, and manganese from municipal water supplies (2, 3, 4, 5). This process takes advantage of the low solubilities of calcium and magnesium compounds and removes these precipitated cations by sedimentation. Calcium is precipitated as calcium carbonate by increasing the carbonate concentration in the water. Magnesium is precipitated as magnesium hydroxide by increasing the hydroxide concentration. Lime and soda ash are the chemicals most often used to bring about these chemical reactions. Precipitation is responsible for iron and manganese removal to levels within drinking water standards or industrial quality.

The total dissolved solids in the water, however, are not greatly affected by this process. Calcium and magnesium ions are replaced by sodium ions while the sulfate concentration remains constant. The choice of this process depends on the divalent cation concentrations in the raw AMD and the concentrations desired in the effluent water.

For the application of lime-soda softening to acid mine drainage, the first four unit processes are the same as for conventional lime neutralization (6). These are equalization, neutralization, iron oxidation, and solids removal, as illustrated in Figure 11-1. Then, the effluent from the solids removal unit enters a flash mix tank for chemical addition, the first softening

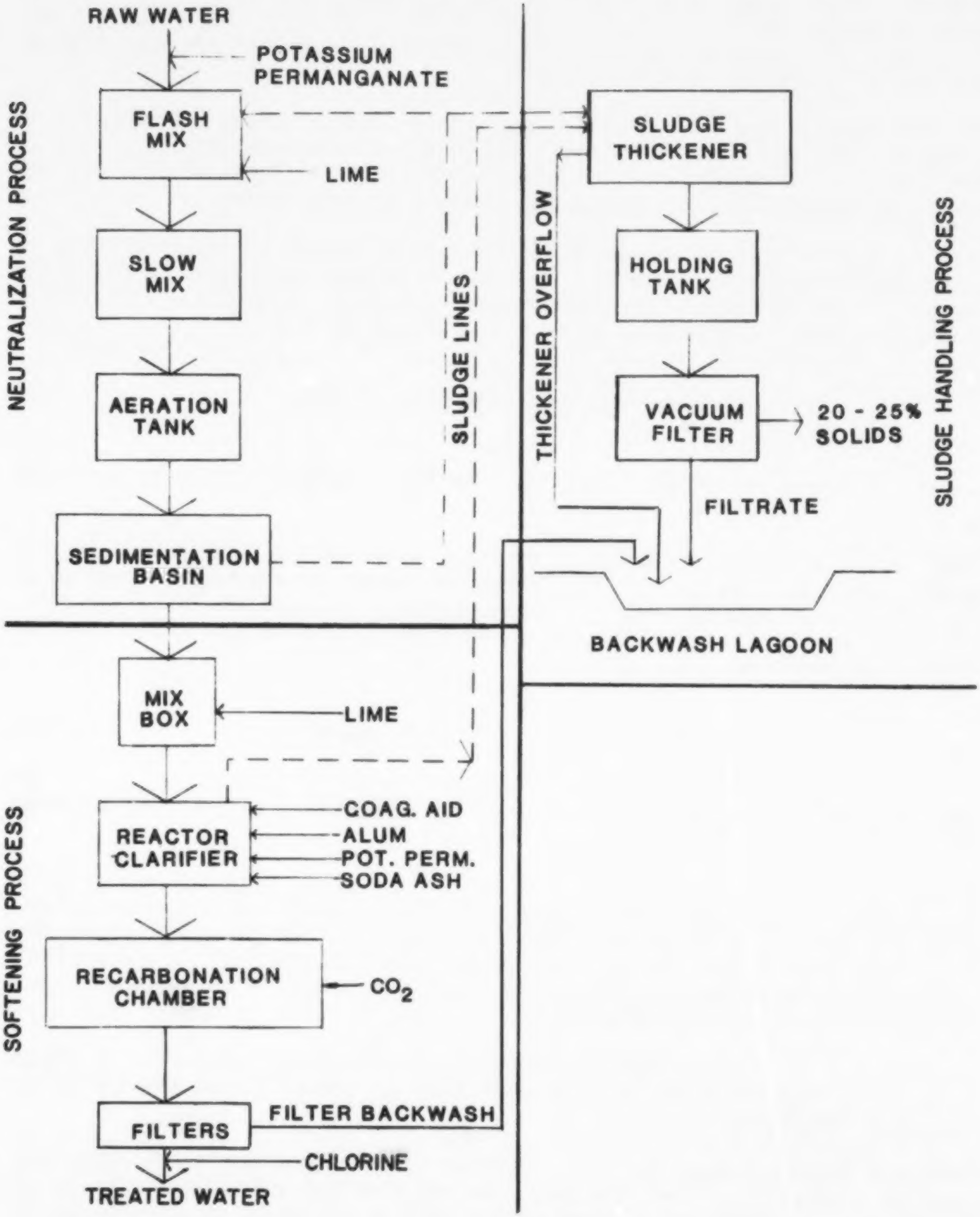
RAW WATER
POTASSIUM PERMANGANATE
FLASH MIX
LIME
SLOW MIX
AERATION TANK
SEDIMENTATION BASIN
NEUTRALIZATION PROCESS
SLUDGE LINES
THICKENER OVERFLOW
SLUDGE THICKENER
HOLDING TANK
VACUUM FILTER
20 - 25% SOLIDS
FILTRATE
SLUDGE HANDLING PROCESS
BACKWASH LAGOON
MIX BOX
LIME
REACTOR CLARIFIER
COAG. AID
ALUM
POT. PERM.
SODA ASH
RECARBONATION CHAMBER
CO_2
FILTERS
FILTER BACKWASH
CHLORINE
TREATED WATER
SOFTENING PROCESS

Figure 11-1. Unit processes of lime-soda softening.

process. This step is followed by the softening reaction (flocculation) tank, settling basins, a recarbonation chamber, filters, and chlorination. Provisions also must be made for sludge recirculation and sludge handling, as well as for filter backwash equipment.

The functions of the neutralization and iron oxidation stages are the same as a typical AMD plant: pH adjustment and iron and manganese removal. Lime is added at this stage for neutralization of the mine drainage acidity and the precipitation of iron, manganese, and aluminum as hydroxides.

Lime is again added to the sedimentation basin effluent as the first step in the softening process. Lime is required for further manganese and magnesium removal, both of which are precipitated as their respective hydroxides. These reactions are expressed by Equations 34 and 35 (6):

$$MgSO_4 + Ca(OH)_2 \rightarrow Mg(OH)_2 + CaSO_4 \tag{34}$$

$$MnSO_4 + Ca(OH)_2 \rightarrow Mn(OH)_2 + CaSO_4 \tag{35}$$

Free carbon dioxide and carbonate hardness also exert a lime demand in this stage. Soda ash (Na_2CO_3) is then added to remove noncarbonate calcium hardness or calcium sulfate ($CaSO_4$), as illustrated in Equations 34, 35, and 36. The calcium ion is precipitated as calcium carbonate while the sulfate remains in solution.

$$CaSO_4 + Na_2CO_3 \rightarrow CaCO_3 + Na_2SO_4 \tag{36}$$

The pH must be maintained at 9.5 or higher for this precipitation to occur (6).

A lime-soda softening plant in Altoona, Pa., was built by the Pennsylvania Department of Environmental Resources for the purpose of treating streams affected by mine drainage to augment the supply of drinking water for that city (see Table 11-1). It is important to note that the sulfate and specific conductance concentrations in the raw water are low, thus qualifying it as a potential source of potable water.

A schematic of the treatment plant process was shown previously in Figure 11-1. The finished water quality of this plant generally met the EPA Drinking Water Standards (6).

During a 5-month study, several important observations concerning this process were noted (6). The most important was that the softening system worked well only when the reaction zone solids were in the range of 10%-15% settleable solids by volume. A failure to build up the solids in the reactor-clarifier unit resulted in an increased effluent hardness. Also, a minimum pH of 11.0 and minimum temperature of 12°C (54°F) were required for favorable

TABLE 11-1

TYPICAL BLENDED RAW WATER CHARACTERISTICS, LIME-SODA SOFTENING PLANT, ALTOON, PA. KITTANNING RUN:IMPOUNDING DAM = 1:1.33

Parameter	Value
pH	3.0
Acidity, mg/l as $CaCO_3$	170
Calcium, mg/l	28
Magnesium, mg/l	18
Iron, mg/l	17
Manganese, mg/l	4.5
Aluminum, mg/l	13
Sodium, mg/l	1.6
Hardness, mg/l as $CaCO_3$	260
Sulfates, mg/l	270
Specific conductance, μmhos/cm	820

softening reactions.

The softener unit (solids contact clarifier) required sludge recirculation to the reaction zone where raw water and chemicals were added. A buildup of reaction zone solids (10%-15% settleable solids by volume in 15 minutes) was a prerequisite for proper operation according to the manufacturer's recommendation. Only during 3 weeks of the study was it possible to obtain this type of operation. The reason for failure of the unit to build up solids was not understood and resulted in disenchantment among the experimenters. The minimum attainable hardness during the study was 120 mg/l $CaCO_3$, regardless of the soda ash dosage. The authors of the study recommended against the use of the particular solids contact clarifier installed in the Altoona plant.

The most effect removal of both manganese and magnesium in the neutralization stage occurred at a pH of 11.0. Provisions for the addition of potassium permanganate ($KMnO_4$) prior to neutralization were made to insure complete manganese oxidation when the pH was less than 11.0. The filters removed significant amounts of iron and manganese, enabling the effluent to meet the drinking water standards of 0.3 mg/l iron and 0.05 mg/l manganese.

The reported costs in 1975 for this treatment process were $\$0.030/m^3$ ($\$0.114/$1,000 gal) for the neutralization in the first stage of the system. The raw water averaged 170 mg/l of acidity and 17 mg/l of total iron. For the overall system operation, costs ranged from $\$0.098/m^3$($0.371/1,000 gal) for water with 120 mg/l of hardness to $\$0.088/m^3$ ($0.333/1,000 gal) for 200 mg/l hardness. These operating costs include power, chemicals, personnel, maintenance supplies, and depreciation.

11.3 Alumina-Lime-Soda Process

The alumina-lime-soda process was originally developed as a method of desalinating brackish water to produce high-quality potable water. It is specifically suited for waters in which the principal source of salinity is sulfate. It is capable of removing heavy metals and hardness as well. Economic considerations indicate the process is most useful for treating AMD with sulfate concentrations between 400 and 1,200 mg/l. The practical lower limit for this process is 100 mg/l, where sulfate removal is not economically feasible. The drinking water standards limit sulfate concentration to 250 mg/l.

The alumina-lime-soda process is divided into two stages, as illustrated in Figure 11-2. The raw AMD is split into two streams, the larger of which is treated with lime and sodium aluminate ($NaAlO_2$) (Stage I). The ratio of AMD treated in Stage I to that bypassed to Stage II depends on the sulfate concentration desired in the total plant effluent. The effluent from Stage I will contain approximately 100 mg/l sulfate, while the sulfate concentration bypassed to Stage II remains constant. The sulfate concentration in the final blended flow can be calculated by a simple mass balance. Stage I effluent is mixed with the smaller AMD stream while carbon dioxide is added for pH control (Stage II). Both stages produce solids, which are removed by filtration.

The key process reactions occur in Stage I. The sodium aluminate and lime neutralize the raw acidity, precipitate the heavy metals and magnesium, and remove calcium sulfate. Pilot studies conducted at the Commonwealth of Pennsylvania, Acid Mine Drainage Research Facility at Hollywood, Pa., indicate that maximum sulfate removal occurs when the Stage I pH is held at 12.0 and the alkalinity level is 600 mg/l as $CaCO_3$. Values less than these result in ineffective sulfate removal. Values greater than these will not effectively remove more sulfate, resulting in wasted lime and increased operating costs.

Sulfate is removed by the sodium aluminate, and to a lesser extent, by the iron and aluminum present in the raw acid mine drainage. The reaction between the sodium aluminate, lime, and AMD produces insoluble calcium sulfoaluminates ($3CaSO_4 \cdot Al_2O_3 \cdot 3CaO \cdot x\ H_2O$ and $CaSO_4 \cdot Al_2O_3 \cdot 3CaO \cdot x\ H_2O$). The latter form seems to dominate the resulting sludge. Each mole of sodium aluminate removes 1 mol of sulfate. The iron present in the AMD forms insoluble calcium sulfoferrites. The lime stabilizes the precipitates as well as

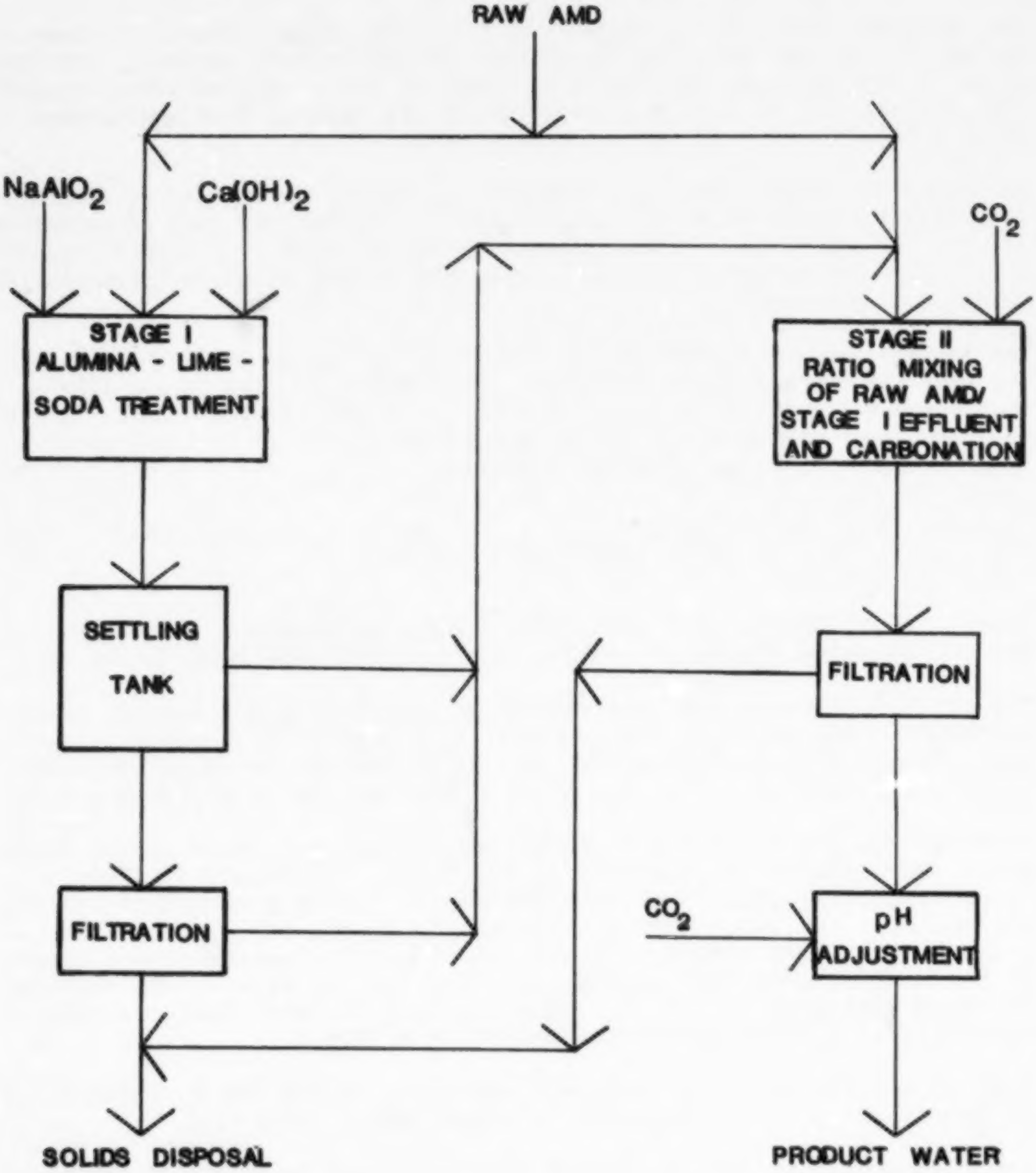
RAW AMD
$NaAlO_2$
$Ca(OH)_2$
CO_2
STAGE I ALUMINA - LIME - SODA TREATMENT
STAGE II RATIO MIXING OF RAW AMD/ STAGE I EFFLUENT AND CARBONATION
SETTLING TANK
FILTRATION
FILTRATION
CO_2
pH ADJUSTMENT
SOLIDS DISPOSAL
PRODUCT WATER

Figure 11-2. Stages of the alumina-lime-soda process.

adding low-cost alkalinity to maintain the reaction pH of 12.0.

The Hollywood experiments found that a minimum Stage I retention time of 80-100 minutes was necessary to provide maximum sulfate removal. The combined factors of high retention time, high pH, and a reaction vessel open to the air resulted in complete oxidation of the ferrous iron and manganese, which eliminated the need for a separate oxidation unit.

Following the Stage I reaction, precipitating solids are removed in a mechanical settling unit. The resulting sludge is usually less than 2% solids by weight, and should be further concentrated by pressure or vacuum filtration to recover the softened filtrate. Dewatered sludge with 10%-12% solids by weight was demonstrated.

The Stage I water contains excess lime, which is removed in Stage II by mixing with the smaller stream of raw acid mine drainage. This excess lime will neutralize the acid in the raw mine drainage stream. Carbon dioxide is also added to lower the pH to 10.3, the minimum solubility of calcium carbonate. Calcium carbonate is formed by the reaction

$$2OH^- + Ca^{+2} + CO_2 \rightarrow CaCO_3 + H_2O \qquad (37)$$

Excess carbon dioxide will redissolve calcium to form calcium bicarbonate. Too little carbon dioxide will leave free hydroxide alkalinity in the water.

The calcium carbonate and metal hydroxide precipitates are removed by sand filtration. The resulting filtrate will have a pH of 10.3 and will contain about 35 mg/l of dissolved calcium carbonate, its minimum solubility. Additional carbonation will drop the pH to a value acceptable for potable water.

Costs for the alumina-lime-soda process depend upon the cost of sodium aluminate. This chemical is available in two forms: the commercially available "dry" form, and the "calcined" form produced by heating a mixture of soda ash and bauxite. The latter form is less expensive, but requires the installation of an aluminate slaker. The quantity of sodium aluminate needed depends on the sulfate, iron, and aluminum concentrations in the acid mine drainage. Increased concentrations of the latter two tend to lower aluminate requirements because of their ability to precipitate sulfate.

Construction and operating costs were estimated for AMD treatment facilities of three sizes in 1975. The costs are summarized in Table 11-2.

The chemical costs include sodium aluminate made by a bauxite-soda ash slaking plant. These cost estimates should be doubled if a "dry" aluminate plant is to be built. Depending on the quantity delivered, the March 1978 cost of dry sodium aluminate is \$545-\$875/kkg (\$494-\$794/ton). The sodium aluminate cost at the time of this study was \$303/kkg (\$275/dry ton), so the total operating costs have increased by approximately 50%.

TABLE 11-2

ESTIMATED COSTS OF THE ALUMINA-LIME-SODA PROCESS (1)

	1,893 m^3/d (0.5 Mgal/d)	3,785 m^3/d (1.0 Mgal/d)	18,925 m^3/d (5.0 Mgal/d
Total construction cost (excluding engineering and legal costs)	\$352,000	\$516,000	\$1,382,000
Operation and maintenance costs (excluding depreciation)			
(\$/m^3)	0.27	0.24	0.21
(\$1,000 gal)	1.04	0.92	0.79

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CHAPTER 12

CALCULATIONS AND PROCEDURES FOR DESIGN OF A MINE DRAINAGE TREATMENT PLANT

12.1 Introduction

This chapter will outline a procedure for the design of a treatment process and related equipment for a mine drainage treatment plant. The purpose of this outline is to give the designer insight into the evaluation of possible alternatives and justification for choosing a certain method or process.

12.2 Design Example I

This particular mine drainage plant will have a daily average flow of 2,880 m^3/d (0.76 Mgal/d), and a raw water quality as shown in Table 12-1.

Several assumptions concerning the operation of this treatment plant have been made to aid in its design. The plant is designed for continuous operation with as little operator supervision as possible. Therefore, the design will be automated as much as possible, and have gravity flow between processes. The processes requiring design are flow equalization, neutralization, aeration, solids separation, and sludge disposal.

12.2.1 Equalization Basin Design (see Chapter 2)

For this design situation, it is assumed that underground storage is available, but it might not be adequate during high flow periods in the spring. Therefore, an equalization basin of 2 days storage volume will be sized. The basin will operate at 25% capacity.

12.2.1.1 Sizing

The equalization volume is

$$\text{Volume} = \frac{2{,}880\ m^3}{d} \times 2\ d = 5{,}760\ m^3\ (203{,}400\ ft^3)$$

TABLE 12-1

RAW WATER QUALITY AND EFFLUENT LIMITATIONS

Parameter	Raw Water	Effluent Limitations 30 Consecutive Day Average	Effluent Limitations Maximum Daily
pH	3.1	6-9	6-9
Net alkalinity, mg/l as $CaCO_3$	-600	>acidity	>acidity
Sulfate, mg/l	1,500	--	--
Suspended solids, mg/l	55	35	70
Iron (total), mg/l	100	3.5	7.0
Iron (ferrous), mg/l	95	--	--
Manganese, mg/l	3	2.0	4.0
Nickel, mg/l	1	--	--
Zinc, mg/l	1	--	--
Aluminum, mg/l	30	--	--

Assuming a 3-m (10-ft) water depth with a 1-m (3-ft) freeboard and choosing to make a square equalization basin, the approximate water surface area is

$$\text{Approximate surface area} = \frac{5{,}760 \text{ m}^3}{3 \text{ m}} = 1{,}920 \text{ m}^2 \text{ } (20{,}670 \text{ ft}^2)$$

The equalization basin will be a square earthen basin with 2½:1 inside slopes and 3:1 outside slopes. Top berm width will be 4.6 m (15 ft) and the inside berm-to-berm length 36.6 m (120 ft). These dimensions provide the required volume, as shown by the average end area calculation.

$$\text{Water Volume} = \frac{(A_1 + A_2)}{2} H = \frac{(36.6)^2 + (51.6)^2}{2} (3) = 6{,}004 \text{ m}^3 \text{ } (211{,}941 \text{ ft}^3)$$

where A_1 and A_2 are the bottom and top water surface areas, respectively, and

H is the water depth

The following calculations show how the cut and fill volumes are balanced to determine the height of the dikes above ground level (h). This assumes the proposed site is relatively flat (see Figure 12-1).

$$\text{Cut Volume} \quad = \quad \text{Fill Volume}$$

$$\frac{(36.6)^2 + (56.4 - 5h)^2}{2}(4 - h) = \frac{4.6 + (4.6 + 5.5h)}{2}(h)(4(36.6) + 4(14.5 + 3h))$$

$$(4{,}521 - 564h + 25h^2)(4 - h)(\tfrac{1}{2}) = (\tfrac{1}{2})(h)(9.1 + 5.5h)(204 + 12h)$$

$$17{,}903 - 2{,}233h + 99h^2 - 4{,}521h + 564h^2 - 25h^3 = 1{,}865h + 1{,}232h^2 + 66h^3$$

$$8{,}619h + 569h^2 + 91h^3 = 17{,}903$$

This equation can be solved for the height of the dikes above ground level (h) by trial and error. Values of h are substituted into the left side of the equation until the solution is reasonably close to the desired value.

h	f (h)
1.52 m (5 ft)	14,779
1.83 m (6 ft)	18,236

Cut Volume = 3,825 m^3 (5,000 yd^3)

Fill Volume = 3,980 m^3 (5,200 yd^3)

If the proposed site is level, 2.1 m (7 ft) of excavation will provide the fill necessary to build the dikes 1.8 m (6 ft) above ground level, giving a total height of 4 m (13 ft).

The total pond area is

$$\text{Pond area} = (76.5\ \text{m})^2 = 5{,}850\ \text{m}^2\ (63{,}000\ \text{ft}^2) = 0.6\ \text{ha}\ (1.5\ \text{ac})$$

12.2.2 Theoretical Lime Requirement (see Chapter 3)

Assume a 70% hydrated lime efficiency. The net alkalinity, determined by analysis, is -600 mg/l as CaCO . The theoretical lime requirement from Table 1-2 is as follows:

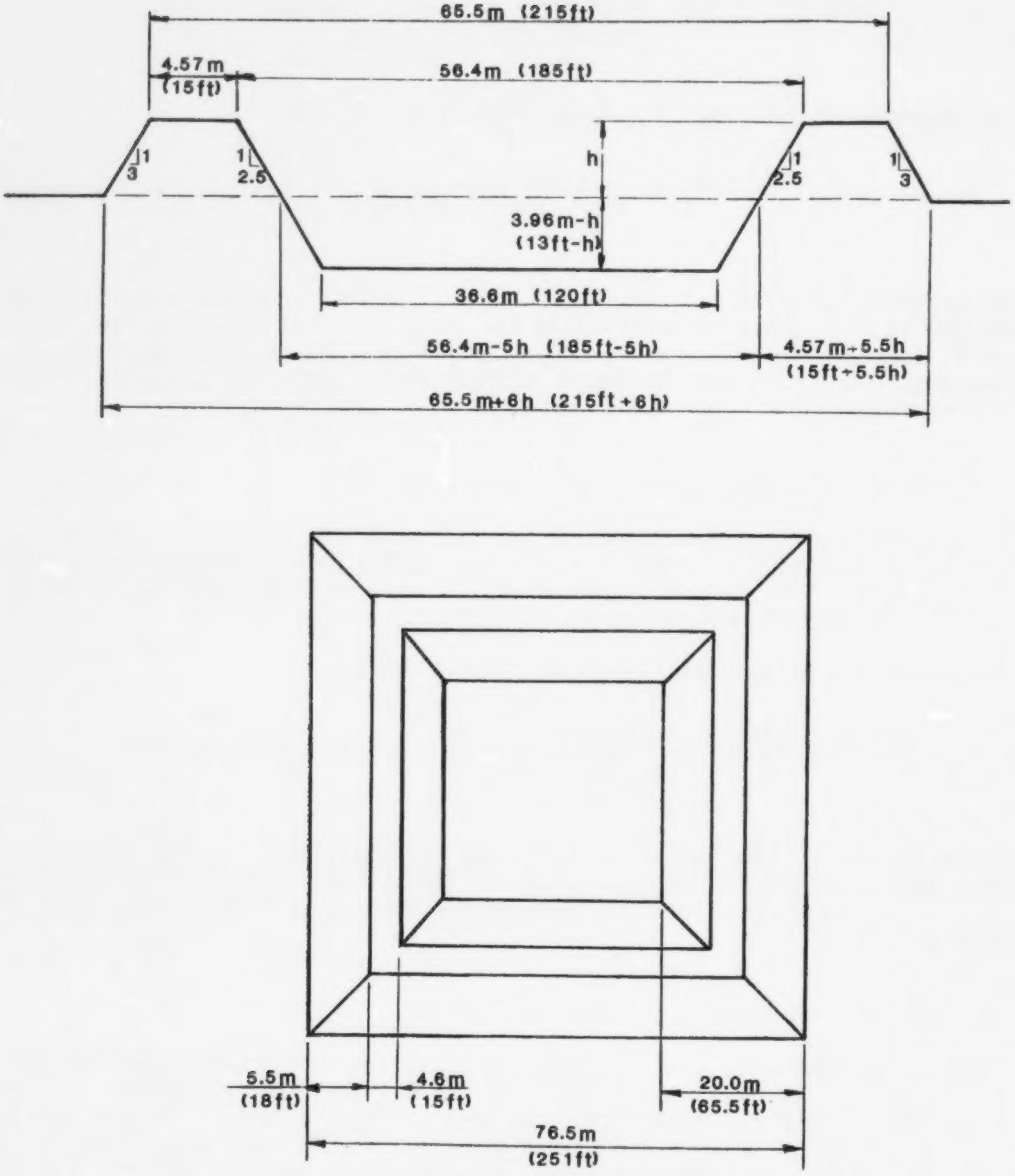
65.5m (215ft)
4.57m (15ft)
56.4m (185ft)
1
3
1
2.5
h
1
2.5
1
3
3.96m-h (13ft-h)
36.6m (120ft)
56.4m-5h (185ft-5h)
4.57m+5.5h (15ft+5.5h)
65.5m+6h (215ft+6h)
5.5m (18ft)
4.6m (15ft)
20.0m (65.5ft)
76.5m (251ft)

Figure 12-1. Equalization basin, Design Example I.

$$600\ \text{mg/l}\ CaCO_3 \times \frac{1\ \text{mol}\ Ca(OH)_2}{1.35\ \text{mol}\ CaCO_3} \times \frac{1}{0.7} = 635\ \text{mg/l}\ Ca(OH)_2$$

The theoretical daily hydrated lime requirement is

$$635\ \text{mg}\ Ca(OH)_2 \times \frac{2{,}880\ m^3}{d} \times \frac{1{,}000\ l}{m^3} \times \frac{1\ kg}{10^6\ mg} = 1{,}830\ \text{kg/d}\ (4{,}025\ \text{lb/d})$$

An operating pH of 8.5 is desired. The theoretical lime requirement, based on the acidity analysis, will raise the pH to 8.3. It is estimated that a small increase in lime must be added to maintain the desired operating pH. Any significantly higher pH would require that a titration curve be established and the lime requirements established on that basis.

12.2.3 Lime Requirement from Treatability Test

A treatability test was conducted according to the procedure outlined in Chapter 6, Section 6.3. This is the better method of determining actual lime requirements. The test indicated that 627 mg/l of lime would be needed to achieve the desired neutralization pH of 8.5.

Therefore, the actual daily lime requirement equals

$$\frac{627\ mg}{l} \times \frac{2{,}880\ m^3}{d} \times \frac{1{,}000\ l}{m^3} \times \frac{1\ kg}{10^6\ mg} = 1{,}800\ \text{kg/d}\ (4{,}000\ \text{lb/d})$$

Since the daily lime usage of 1.8 kkg/d (2 tons/d) is low, the designer's best choice would be to employ a hydrated lime system. A quicklime operation with slaker requires a capital investment and maintenance cost that would be difficult to justify.

12.2.4 Lime Silo Sizing (see Chapter 3)

Considering that most mine drainage plants are in rural areas, the designer should provide at least seven days lime storage capacity in a silo. In this case, the minimum silo capacity should be 12.7 kkg (14 tons). To take advantage of pneumatic bulk delivery (minimum 18.1 kkg (20 tons) per delivery), a silo larger than 18.1 kkg (20 tons) is necessary. The cost per unit volume of silos is perhaps the least expensive item in the entire plant. Thus, the designer can provide excess capacity at relatively low costs.

A 27.2-kkg (30-ton) silo is chosen. This will permit bulk delivery, allow a 9.1-kkg (10-ton) floating freeboard for operation, and give the operator a 5-day leeway for delivery.

Assuming a hydrated lime density of 481 kg/m^3 (30 lb/ft^3), the silo volume required to store 27.2 kkg (30 tons) is

$$\text{Silo volume} = \frac{27{,}200 \text{ kg}}{481 \text{ kg/m}^3} = 56.5 \text{ m}^3 \text{ (2,000 ft}^3\text{)}$$

The silo specifications are

diameter = 3.05 m (10 ft)
sidewall height = 7.3 m (24 ft)
60° hopper bottom

12.2.4.1 Bin Activator (Vibratory)

A vibratory bin activator is highly recommended for hydrated lime silos. Bin activators are sized as one-half the silo diameter for silos up to 6.1 m (20 ft) in diameter, and one-third the diameter for larger silos. Therefore, a 1.52-m (5-ft) vibratory bin activator should be used.

12.2.4.2 Other Silo Equipment Required

1. Dust collector, whose size varies with model;
2. 10.16-cm (4-in) fill line with long radius turn and quick-connect coupling;
3. bin level indicators (side ports);
4. OSHA side-mounted ladder and a hand railing atop the silo;
5. foundation.

12.2.4.3 Silo Slide Valve

It is necessary to have a silo slide valve on the bin activator bottom. This enables the operator to close the silo at any time, and is extremely important when the feeder malfunctions or when an empty silo is being filled. When filling an empty silo, this valve must be closed. Otherwise, lime will be blown throughout the feeder area.

12.2.5 Lime Feeder (see Chapter 3)

The average hydrated lime feed rate is as follows:

$$\frac{1{,}800\ \text{kg/d}}{24\ \text{hr/d}} = 75\ \text{kg/hr}\ (167\ \text{lb/hr})$$

The volumetric feed rate is

$$\frac{75\ \text{kg/hr}}{480\ \text{kg/m}^3} = 0.16\ \text{m}^3\text{/hr}\ (5.52\ \text{ft}^3\text{/hr})$$

The designer has a number of feeder types from which to choose; the choice can be a matter of personal preference. A variable screw feeder with a delivery range of 0.065-0.65 m^3/hr (2.3-23 ft^3/hr) is chosen.

12.2.6 Optional Designs for Lime Feeding

Since the lime requirement for this particular plant is low, the designer has methods available other than a conventional lime slurry feed system. A dry feed system or a volume slurrying system are viable alternatives.

12.2.6.1 Dry Feed System

There are various opinions on the feeding of dry lime directly into the raw water. There exists a cutoff point, yet to be determined, where dry lime feed has prominent disadvantages. This stems from the low solubility of lime. In this case, however, where the lime requirement is only 0.625 kg/1,000 l (5.26 lb/1,000 gal), a dry lime system is entirely feasible.

The designer, as a modification to the previous design, can eliminate the slurry feed system, but should provide a longer reaction time in the flash mixer.

12.2.6.2 Volume Slurrying

Volume slurrying involves blowing the hydrated lime into a large closed-top tank equipped with a mixer and large dust collector. The tank contains a specific amount of water that depends upon the slurry concentration desired. The lime slurries as it enters the tank and is continuously agitated. The slurry tank must have the capacity to handle 18.1 kkg (20 tons) of dry lime put in solution. If a 15% slurry is desired, a minimum tank capacity of 122 m^3 (32,260 gal) is required to accept a full pneumatic truck load. To maintain continuous treatment, at least 4.54 kkg (5 tons) or 30.3 m^3 (8,000 gal) capacity would have to be added, increasing the minimum slurry tank volume to 151.4 m^3 (40,000 gal).

12.2.6.3 Lime Slurry System

A 10% (by weight) lime slurry concentration is initially chosen as the feed (see Table 3-1). The slurry weight is 1.1 kg/l (8.8 lb/gal). The lime ($Ca(OH)_2$) concentration is 0.10 kg/l (0.83 lb/gal). The required slurry makeup rate is

$$\frac{75 \text{ kg/hr}}{0.10 \text{ kg/l}} \times \frac{1 \text{ hr}}{60 \text{ min}} = 12.5 \text{ l/min } (3.3 \text{ gal/min})$$

However, the slurry system makeup rate should be at least two times that required. Therefore, a 25 l/min (6.6 gal/min) makeup rate should be used, with a water-to-lime weight ratio of 9.7:1. The lime feed should be

$$25 \text{ l/min} \times \frac{0.1 \text{ kg}}{\text{l}} Ca(OH)_2 \times \frac{60 \text{ min}}{\text{hr}} = 150 \text{ kg } Ca(OH)_2\text{/hr } (330 \text{ lb/hr})$$

$$= 0.3 \text{ m}^3\text{/hr } (11 \text{ ft}^3\text{/hr})$$

$$= 2.5 \text{ kg/min } (5.5 \text{ lb/min})$$

The water feed should be

$$2.5 \text{ kg/min} \times 9.7 = 24.2 \text{ kg } H_2O\text{/min } (53.4 \text{ lb/min})$$

$$= 24.2 \text{ l } H_2O\text{/min } (6.4 \text{ gal/min})$$

12.2.6.4 Slurry Tank

The slurry tank will be designed to provide 20 minutes detention time when half full. The tank will provide slurry preparation, storage, and stabilization.

The required tank volume is

$$\text{Volume} = 40 \text{ min} \times 12.5 \text{ l/min} \times \frac{1 \text{ m}^3}{1{,}000 \text{ l}} = 0.5 \text{ m}^3 \; (17.7 \text{ ft}^3)$$

Design a circular tank with the diameter equal to the height.

$$\text{Volume} = 0.785 \, (D^2) \, (H) \qquad \text{Set } D = H$$

$$0.5 = 0.785\ (D^3)$$

$$D = 0.86\ \text{m}\ (2.8\ \text{ft})$$

Make the tank diameter 0.9 m (3 ft) and the overall height 1.22 m (4 ft), which includes 0.3 m (1 ft) freeboard. The total working volume is 592 l (160 gal). The lime feeder and water will be turned on when the low level is reached, and turned off at high level.

The control level positions, with respect to tank bottom, will be as follows:

Low level (turns feeder and water on) at 0.46 m (1.5 ft)
High level (turns feeder and water off) at 0.91 m (3 ft)
Emergency high level (shutdown of system) at 1.07 m (3.5 ft)

12.2.6.5 Slurry Mixer

A standard rule of thumb to size the mixer is 0.2 kW/m^3 (1 hp/1,000 gal). This will give the designer a good approximation of the mixer size until the final sizing can be determined with the help of the vendor.

Since the slurry tank volume is only 0.592 m^3 (160 gal), the required horsepower equals

$$0.592\ \text{m}^3 \times (0.2\ \text{kW/m}^3) = 0.11\ \text{kW}\ (0.15\ \text{hp})$$

Mixer motors this size operate at 30%-50% efficiency. Therefore, a 0.37 kW (0.5 hp), angular, side-mounted propeller mixer should be provided. The propeller shaft should extend below the low-level control to insure that the contents will be mixed at all times.

12.2.6.6 Slurry Feed System

One variable-speed, volumetric slurry feeder should be provided, with an operating range of 0.314-113.6 l/min (0.083-30 gal/min) and a 0.19-kW (0.25-hp) feeder motor. This variable speed feeder is controlled by a pH monitor placed immediately after the flash mixing tank. The pH probes should not be placed inside the flash mix tank, but as near to the inlet of the next (aeration or settling) unit as possible.

12.2.6.7 Piping and Feed Pump

Ideally, the slurry line velocity within the slurry feed loop should be in the range of 1.0-1.2 m/s (3-4 ft/s). The pipe diameter required to maintain this velocity at a flow rate of 12.5 l/min (3.3 gal/min) is calculated as follows:

$$\text{Pipe cross-sectional area} = \frac{\text{Flow}}{\text{Velocity}} = \frac{12.5\ \text{l/min}}{1.2\ \text{m/s}} \times \frac{\text{min}}{60\ \text{s}} \times \frac{1{,}000\ \text{cm}^3}{1} \times \frac{\text{m}}{100\ \text{cm}}$$

$$= 1.71\ \text{cm}^2\ (0.266\ \text{in}^2)$$

This cross-sectional area would indicate a commercial pipe of 1.27-cm (0.5-in) diameter; however, it is also recommended that any slurry line not be less than 2.5 cm (1 in) in diameter. Assuming a 2.5-cm (1-in) minimum pipe diameter for this slurry line, and use of head loss tables for standard water pipe, a flow of 37.9 l/min (10 gal/min) will produce a velocity of about 1.1 m/s (3.7 ft/s), with a head loss of 12 m/100 m (12 ft/100 ft). These are acceptable design values.

In some cases, revisions to the sizing of the slurry tank, its mixer, or other related equipment may be necessary. In this case, the slurry is to be recirculated from the slurry makeup tank to the volumetric feeder. Excess flow is returned to the slurry tank and a pH system controls the feed from the volumetric feeder to the flash mix tank.

The head loss in a 2.54-cm (1-in) diameter pipe is approximately 12 m/100 m (12 ft/100 ft). If it is assumed that the length of the slurry loop is 7.62 m (25 ft) and the static head is 3.0 m (10 ft), the total pump head is

$$\text{Total head} = 3.0 + 7.62 \times \frac{12}{100} = 3.91\ \text{m}\ (12.8\ \text{ft})$$

The slurry pump power needed is

$$\text{kW} = \frac{rQH}{101.97}$$

where r = slurry density (kg/m^3)

Q = slurry flow (m^3/s)

H = head (m)

$$\text{kW} = \frac{1{,}060\ \text{kg/m}^3 \times 0.0024\ \text{m}^3\text{/s} \times 3.91\ \text{m}}{101.97}$$

$$\text{kW} = 0.098\ \text{kW}\ (0.13\ \text{hp})$$

According to vendor catalogs, pumps in this range operate at 35% efficiency, which is very low. Thus, the required pump size would be as follows:

$$\frac{0.098\ \text{kW}}{0.35} = 0.28\ \text{kW}\ (0.037\ \text{hp})$$

A slurry pump with a 0.37-kW (0.50-hp) motor with a 2.54-cm (1-in) discharge and 3.18-cm (1¼-in) suction should be selected.

12.2.7 Flash Mix Tank (see Chapter 4)

Assume an effective detention time of 5 minutes. The volume required must include both the AMD flow and the slurry flow.

$$\text{Vol} = 5\ \text{min} \times \left(\left(2{,}880\ \text{m}^3/\text{d} \times \frac{1\ \text{d}}{1{,}440\ \text{min}} \times \frac{1{,}000\ \text{l}}{1\ \text{m}^3}\right) + 12.5\ \text{l/min}\right)$$

$$= 10{,}063\ \text{l}\ (2{,}660\ \text{gal})$$

Design a circular tank with H:D approximately equal to 1.

$$\text{Volume} = 0.785\ (D^2)\ (H)$$

$$10.06\ \text{m}^3 = 0.0785\ D^3$$

$$D = 2.34\ \text{m}\ (7.68\ \text{ft})$$

The final flash mix tank dimensions are

Diameter = 2.44 m (8 ft)

Height = 2.44 m (8 ft) + 0.61 m (2 ft) freeboard = 3.05 m (10 ft)

The inlet and outlet positions are 180° apart.

Inlet - top entry of raw water and slurry pipes with air break.

Outlet - pipe insuring 1.0-1.2 m/s (3-4 ft/s) exit velocity.

The outlet pipe diameter calculations are

$$\text{Area} = \frac{\text{Flow}}{\text{Velocity}}$$

$$A = \frac{(2{,}880\ m^3/d \times \frac{1\ d}{1{,}440\ min}) + (12.5\ l/min \times \frac{m^3}{1{,}000\ l})}{1.0\ m/s \times \frac{60\ s}{min}}$$

$A = 0.034\ m^2$

$D = 0.2\ m$

Use a 20-cm (8-in) pipe. Use a 20-cm (8-in) inverted elbow with a 0.61-m (2-ft) nipple entering at the 2.44-m (8-ft) height.

The tank will require baffling. Place four baffles 90° apart. Design the baffles according to the following specifications:

Baffle width 1/18 x 244 cm = 13.6 cm (5.3 in)

Make the baffle width 15.24 cm (6 in).

Wall clearance = 0.10 x 15.24 cm = 1.42 cm (0.6 in)

Make the minimum clearance 2.54 cm (1 in).

The baffles should end a minimum of 6.35 cm (2.5 in) above the tank bottom. They should extend at least 15.24 cm (6 in) above the static water level.

12.2.8 Aeration Tank (see Chapter 5)

The average ferrous iron concentration is 95 mg/l. The daily ferrous iron loading is

$$\frac{95\ mg}{l} \times \frac{2{,}880\ m^3}{d} \times \frac{1{,}000\ l}{m^3} \times \frac{kg}{10^6\ mg} = 274\ kg\ Fe^{+2}/d\ (603\ lb\ Fe^{+2}d)$$

The theoretical oxygen requirement for iron oxidation is determined by the chemical relationship that

7 kg (15.4 lb) of iron are oxidized by 1 kg (2.2 lb) of oxygen

3.2 kg (7 lb) of iron are oxidized by 0.454 kg (1 lb) of oxygen

$$\frac{274\ kg\ Fe^{+2}/d}{7\ kg\ Fe^{+2}/1\ kg\ O_2} = 39\ kg\ O_2/d\ (86\ lb\ O_2/d) = 1.63\ kg/hr\ (3.59\ lb/hr)$$

The approximate kW (hp) requirements, using 2.13 kg O_2/kW-hr (3.5 lb O_2/hp-hr) oxygen transfer rate, can be determined. A 0.75-kW (1-hp) aerator is initially sized, although mixing requirements must still be considered and are of the utmost importance.

Detention time required in an aeration basin can best be estimated from the iron oxidation-vs.-time curve produced during the treatability study. The slope of this curve given in milligrams per liter per minute can be used to determine the minimum detention time necessary for complete ferrous iron oxidation at the operating pH chosen.

From the treatability test, an 11.25 mg/l/min iron oxidation rate was derived. Assuming complete oxidation in the basin, the detention time required is

$$\text{Detention time} = \frac{95.0 \text{ mg/l}}{11.25 \text{ mg/l/min}} = 8.44 \text{ min}$$

A high degree of short-circuiting can occur in an aeration basin. Therefore, a scale-up or safety factor must be applied. Since this calculated detention period is short, a factor of at least 4.0 (Table 5-2) must be applied. Thus, a minimum 34.0-minute detention time should be provided.

Aeration tank size:

$$\text{Volume} = Q \times t$$

$$= \left(\left(2{,}880 \text{ m}^3/\text{d} \times \frac{1 \text{ d}}{1{,}440 \text{ min}}\right) + \left(12.5 \text{ l/min} \times \frac{\text{m}^3}{1{,}000 \text{ l}}\right)\right) \times 34 \text{ min}$$

$$= 68.4 \text{ m}^3 \ (2{,}400 \text{ ft}^3)$$

where Q = total flow (raw water and slurry)

t = 34 min

Assume a 1.83-m (6-ft) water depth and circular basin.

$$\text{Surface Area} = \frac{68.4 \text{ m}^3}{1.83 \text{ m}^3}$$

$$= 37.4 \text{ m}^2 \ (400 \text{ ft}^2)$$

$$\text{Basin Diameter} = 6.9 \text{ m} \ (20.7 \text{ ft})$$

The designer now selects an aerator that satisfies both oxygen transfer and mixing requirements.

The type of aerator chosen here is a floating, 0.89-kW (1.2-hp) aerator with a working volume of 8.23 m (27 ft) in diameter and 1.83 m (6 ft) in depth, and an oxygen transfer rate of 2.3 kg O_2/kW-hr (3.8 lb O_2/hp-hr).

It is the designer's option either to build the aeration basin from concrete and steel or to use an earthen basin with an erosionproof bottom.

12.2.9 Estimated Sludge Production (see Chapter 7)

The sludge produced by the treatment of this mine water can be estimated as the sum of the metal hydroxides removed in significant concentrations (iron and aluminum), the suspended solids in the mine drainage, and unused lime. This estimation assumes 100% solids removal.

Hydrolysis of ferric iron, assuming complete oxidation of ferrous iron:

$$\begin{array}{ll} Fe^{+3} + 3H_2O & Fe(OH)_3 + 3H^+ \\ 56\text{g/g-mol} & 107 \text{ g/g-mol} \\ 100 \text{ mg/l} & \frac{107}{56}(100 \text{ mg/l}) = 191 \text{ mg/l} \end{array}$$

Aluminum hydrolysis:

$$\begin{array}{ll} Al^{+3} + 3H_2O & Al(OH)_3 + 3H^+ \\ 27 \text{ g/g-mol} & 78 \text{ g/g-mol} \\ 30 \text{ mg/l} & \frac{78}{27}(30 \text{ mg/l}) = 87 \text{ mg/l} \end{array}$$

Theoretical daily solids production:

Ferric hydroxide

$$191 \text{ mg/l} \times \frac{2{,}880 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = 550 \text{ kg/d } (1{,}212 \text{ lb/d})$$

Aluminum hydroxide

$$87 \text{ mg/l} \times \frac{2{,}880 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = 251 \text{ kg/d } (552 \text{ lb/d})$$

Suspended solids

$$55 \text{ mg/l} \times \frac{2{,}880 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = 158 \text{ kg/d } (349 \text{ lb/d})$$

Unused lime (assume 15% lime wastage, which is not unusually high and is due primarily to the insolubility of lime)

1,806 kg/d x 0.15 = 271 kg/d (596 lb/d)

Assuming the sludge withdrawn from the settling basin will be at only 1% solids, the liquid sludge weight is

$$\frac{1{,}230 \text{ kg solids/d}}{0.01} = 123{,}000 \text{ kg sludge/d } (270{,}600 \text{ lb sludge/d})$$

For simplicity, assume the sludge weighs the same as water.

$$\frac{123{,}000 \text{ kg/d}}{1 \text{ kg/l}} = 123{,}000 \text{ l sludge/d } (32{,}497 \text{ gal sludge/d})$$

$$= 123 \text{ m}^3 \text{ sludge/d } (4{,}343 \text{ ft}^3 \text{ sludge/d})$$

The designer can now cross-check this theoretical sludge volume estimation against results found in the treatability study. In this case the settling test gave a final settled volume of 35 ml. The sludge volume then can be calculated as a simple percent.

$$\frac{35 \text{ ml sludge}}{1{,}000 \text{ ml sample}} = 0.035 \text{ or } 3.5\% \text{ of the flow}$$

$$= 0.035 \times 2{,}880 \text{ m}^3\text{/d} = 101 \text{ m}^3\text{/d } (3{,}560 \text{ ft}^3\text{/d})$$

Therefore, the designer should consider the higher sludge volume and design accordingly.

12.2.10 Settling Basin Design

In this case, land is available, so an earthen settling basin is chosen. The design parameters are as follows:

1. detention time equals a minimum of 12 hours clear water storage and a minimum of 3 days sludge storage capacity;
2. equipped with automatic sludge removal device;
3. influent flows equal 2,880 m ^{3}/d (0.76 Mgal/d);
4. 0.91 m (3 ft) freeboard;

5. exterior slopes 3:1;

6. interior slopes 2½:1.

Calculations of basin volume follow:

Three days sludge storage = $123\ m^3/d \times 3\ d = 369\ m^3$ ($13,030\ ft^3$)

$$12 \text{ hours clear water storage} = 2,880\ m^3/d \times \frac{12\ hr}{24\ hr/d}$$

$$= 1,440\ m^3\ (50,900\ ft\ )$$

Total Volume Lime = $1,809\ m^3$ ($63,785\ ft^3$)

Assume a 3.05-m (10-ft) depth using a length-to-width ratio of 3:1 for bottom.

12.2.11 Settling Basin with Hydraulic Sludge Removal System

The designer has decided to use an hydraulic sludge removal system to minimize labor costs associated with manual sludge removal. The manufacturer of these systems has recommended an area of 9.1 m x 18.3 m (30 ft x 60 ft) to be covered by the system. Thus, the pond bottom should have a width of 9.1 m (30 ft) and length of 24.4 m (80 ft) to obtain the necessary volume (see Figure 12-2). The length-to-width ratio is 2.67 and the volume provided is

$$\text{Volume} = \frac{(9.1 \times 24.4) + (24.4 \times 29.6)}{2}\ (3.05) = 1,811\ m^3\ (63,952\ ft^3)$$

which is about equal to the required volume of $1,809\ m^3$ ($63,875\ ft^3$).

Calculations to determine cut and fill volumes, assuming the pond is to be built on level ground, are

$$\text{Cut Volume} = \text{Fill Volume}$$

$$\frac{(24.4)\ (9.1) + (44.2 - 5h)\ (29.0 - 5h)}{2}\ (3.96 - h) =$$
$$\frac{(4.57 + (4.57 + 5.5h)}{2}\ (h)\ ((2)\ (9.1 + 14.5 + 3h) + (2)\ (2.44 + 14.5 + 3h))$$

$$(222 + 1,282 - 145h - 221h - 25h^2)\ (½)\ (3.96 - h) =$$
$$(½)\ (h)\ (9.14 + 5.5h)\ (125 + 12h)$$

$$(1,504 - 366h + 25h^2)\ (3.96 - h) = (h)\ (9.14 + 5.5h)\ (125 + 12h)$$

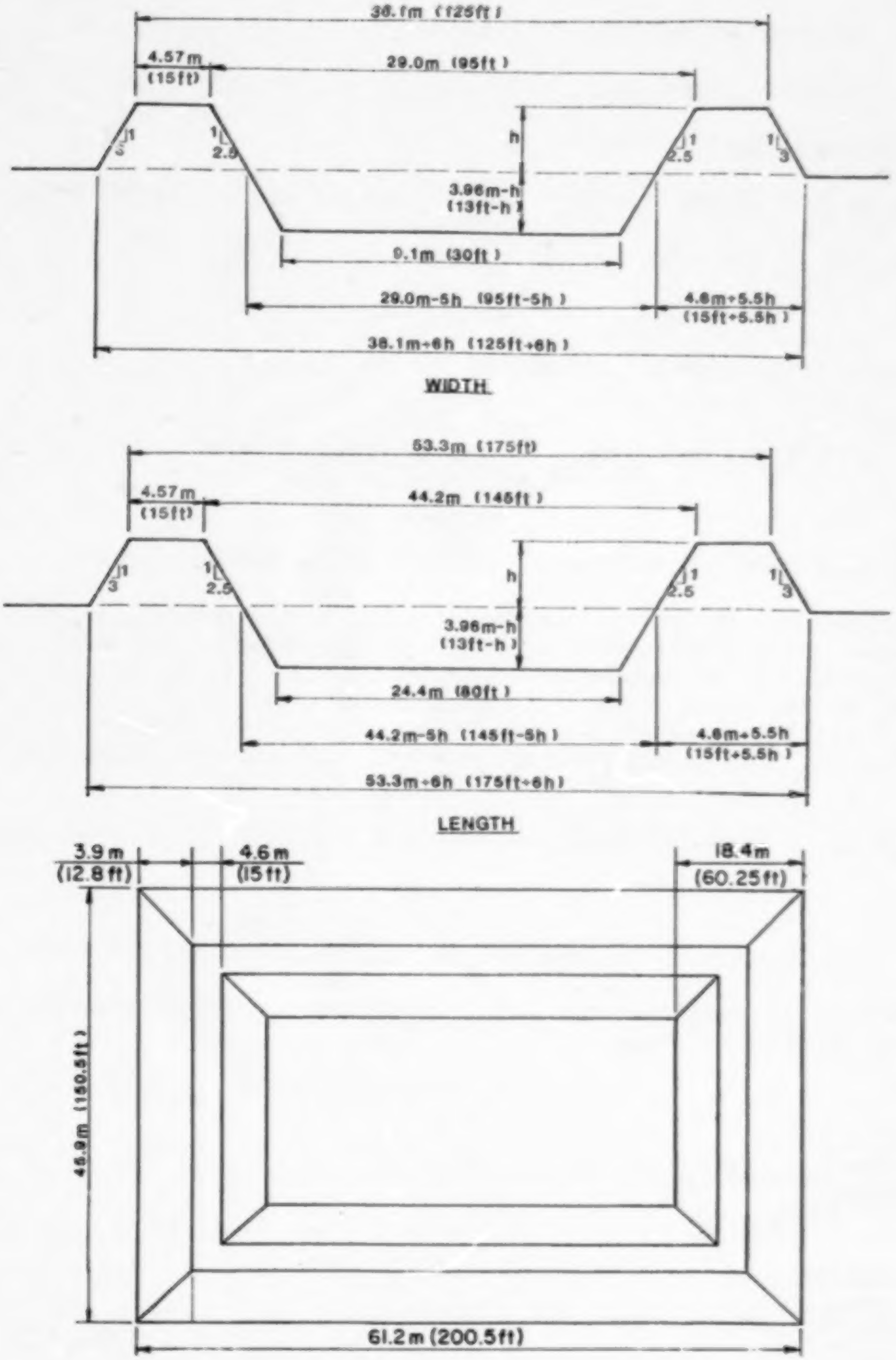
38.1m (125ft)
4.57m (15ft)
29.0m (95ft)
h
1
3
1
2.5
3.96m-h (13ft-h)
9.1m (30ft)
29.0m-5h (95ft-5h)
4.6m+5.5h (15ft+5.5h)
38.1m+6h (125ft+6h)
WIDTH
53.3m (175ft)
4.57m (15ft)
44.2m (145ft)
h
1
3
1
2.5
3.96m-h (13ft-h)
24.4m (80ft)
44.2m-5h (145ft-5h)
4.6m+5.5h (15ft+5.5h)
53.3m+6h (175ft+6h)
LENGTH
3.9m (12.8ft)
4.6m (15ft)
18.4m (60.25ft)
45.9m (150.5ft)
61.2m (200.5ft)

Figure 12-2. Settling basin, Design Example I.

$$(5{,}956 - 1{,}449h + 99h^2 - 1{,}504h + 366h^2 - 25h^3 = 1{,}143h + 688h^2 + 110h^2 + 66h^3)$$

$$4{,}096h + 333h^2 + 91h^3 = 5{,}956$$

This equation can be solved by trial and error.

h	f(h)
1.22 m (4 ft)	5,658
1.37 m (4.5 ft)	6,470
1.30 m (4.25 ft)	6,087

Choose h = 1.30 m (4.25 ft)

Cut Volume = 1,430 m^3 (1,870 yd^3)

Fill Volume = 1,480 m^3 (1,940 yd^3)

Overall Dimensions = 9.1 + 2 (14.5 + (3) (1.30)) = 45.9 m (150.5 ft)

= 24.4 + 2 (14.5 + (3) (1.30)) = 61.2 m (200.5 ft)

Total Basin Area = (45.9 m) (61.2 m) = 2,809 m^2 (3,240 ft^2)

= 0.28 ha (0.69 ac)

Total Water Area = (24.4) (39.6) = 966 m^2 (10,400 ft^2)

= 0.10 ha (0.24 ac)

12.2.12 Sludge Disposal Pond

Based on past experience, the following assumptions have been made for operating a lagoon disposal: a 20-year mine life, and sludge withdrawn at 1% solids for the primary settling basin.

Air-dried sludge can achieve 16%-18% solids in a disposal lagoon, but 12% will be used on a conservative basis. The decanted water is returned to the primary settling pond.

The sludge produced (from treatability test) is as follows:

Sludge produced per day = 123 m^3 (4,343 ft^3) at 1% solids

Final sludge volume at 12% = 10.3 m^3 (362 ft^3)

12.2.12.1 Required Sludge Disposal Pond Volume

$$\frac{10.3 \text{ m}^3}{\text{d}} \times \frac{365 \text{ d}}{\text{yr}} \times 20 \text{ yr} = 75{,}190 \text{ m}^3 \ (2{,}655{,}000 \text{ ft}^3)$$

Provide a 1.52-m (5-ft) clear water depth plus a 0.91-m (3-ft) freeboard at capacity. Assume a pond depth of 10.67 m (35 ft), thus providing an 8.23-m (27-ft) sludge depth.

$$\text{Area required} = \frac{75{,}190 \text{ m}^3}{8.23 \text{ m}} = 9{,}136 \text{ m}^2 \ (98{,}303 \text{ ft}^2)$$

$$= 0.91 \text{ ha } (2.26 \text{ ac})$$

Preliminary pond volume = 95.6 m x 95.6 m x 10.67 m (315 ft x 315 ft x 35 ft)

12.2.12.2 Sludge Pond Layout

Assuming the following parameters, the sludge pond was sized.

Pond inside slope 2½:1   Outside slope 3:1   Berm width = 4.51 m (15 ft)

a. First Trial

Estimate square pond bottom dimensions at 76.2 m x 76.2 m (250 ft x 250 ft):

$$\text{Effective Volume} = \frac{(76.2)^2 + (117.3)^2}{2} (8.23) = 80{,}513 \text{ m}^3 \ (2{,}843{,}274 \text{ ft}^3)$$

which is slightly greater than the required volume, 25,190 m^3 (2,655,000 ft^3).

$$\text{Cut Volume} = \text{Fill Volume}$$

$$\frac{(76.2)^2 + (129 - 5h)^2}{2} (10.7 - h) = \frac{4.6 + (4.6 + 5.5h)}{2} (h)$$

$$(4(129.7 + 5h) + 4(4.6 + 5.5h))$$

$$(22{,}629 - 1{,}297h + 25h^2)(10.7 - h)(\tfrac{1}{2}) = (9.2 + 5.5h)(h)(537 + 2h)(\tfrac{1}{2})$$

$$242{,}130 - 13{,}878h + 268h^2 - 22{,}629h + 1{,}297h^2 - 25h^3 = 4{,}940h + 18.4h^2 + 2{,}954h^2 + 11h^3$$

$$242{,}130 - 41{,}447h - 1{,}407h^2 - 36h^3 = 0$$

$$36h^3 + 1{,}407h^2 + 41{,}447h = 242{,}130$$

This equation can be solved by trial and error.

h	f(h)
4.5 m (14.7 ft)	218,284
5.0 m (16.4 ft)	246,910
4.9 m (16.1 ft)	241,108

$$h = 4.9 \text{ m } (16.1 \text{ ft})$$

$$\begin{aligned}\text{Cut Volume} &= 48{,}933 \text{ m}^3 \ (64{,}726 \text{ yd}^3)\\ \text{Fill Volume} &= 48{,}446 \text{ m}^3 \ (64{,}082 \text{ yd}^3)\end{aligned}$$

The volume to be excavated is equal to the fill needed for the dikes when the depth of excavation is 5.8 m (19.0 ft) and the dikes are raised 4.9 m (16.1 ft) above the existing ground level, making a total depth of 10.7 m (35.1 ft). This assumes level ground at the excavation site.

$$\text{Total area required} = \frac{(161 \text{ m})^2}{10{,}000 \text{ m}^2/\text{ha}} = 2.6 \text{ ha } (6.4 \text{ ac})$$

12.3 Design Example II

This treatment plant is to be designed to treat a flow of 11,520 m^3/d (3.04 Mgal/d) with a raw water quality as shown in Table 12-2.

Several assumptions concerning the operation of this plant have been made to aid in its design. The plant is designed to operate continuously with as much automation as possible, reducing the need for operator supervision to one visit per day. Gravity flow between the unit processes should be provided. These processes should include flow equalization, pH adjustment, aeration, solids separation, and sludge disposal. It is also assumed that limited land area is available, and that final sludge disposal is to a deep mine through a borehole.

12.3.1 Equalization Basin

Because large underground storage is available, only a small equalization basin need be constructed (1 day storage). This basin will be at least 11,520 m^3 (3.04 Mgal) in size.

Assume a 3.0-m (10-ft) water depth and a 0.9-m (3-ft) freeboard above the maximum water level.

TABLE 12-2

RAW WATER AND EFFLUENT QUALITY LIMITATIONS

Parameter	Raw Water	Effluent Limitations 30 Consecutive Day Average	Effluent Limitations Maximum Daily
pH	2.9	6-9	6-9
Acidity	2,200	net acidity	net acidity
Sulfate, mg/l	2,000	--	--
Suspended solids, mg/l	165	35	70
Iron (total), mg/l	480	3.5	7.0
Iron (ferrous), mg/l	440	--	--
Manganese, mg/l	11	2.0	4.0
Nickel, mg/l	1.5	--	--
Zinc, mg/l	3	--	--
Aluminum, mg/l	150	--	--

$$\text{Approximate Water Surface Area Requirement} = \frac{11{,}520\ \text{m}^3}{3\ \text{m}}$$

$$= 3{,}840\ \text{m}^2\ (41{,}330\ \text{ft}^2)$$

Design the equalization basin to be a square earthen basin with 2½:1 inside slopes and 3:1 outside slopes. Make the top berm width 4.6 m (15 ft) and the inside berm-to-berm length 54.9 m (180 ft). These dimensions will give the volume required, as shown by the following calculation:

$$\text{Water Volume} = \frac{(A_1 + A_2)}{2}\,H = \frac{(54.9)^2 + (70.1)^2}{2}\,(3) = 11{,}900\ \text{m}^3\ (420{,}000\ \text{ft}^3)$$

where A_1 and A_2 are the bottom and top water surface areas respectively, and H is the water depth.

The following calculations show how the cut and fill volumes are balanced to determine the proper excavation depth. It is assumed that the land area before excavation is flat (see Figure 12-3).

Cut Volume = Fill Volume

$$\frac{(54.9)^2 + (74.7 - 5h)^2}{2} (3.96 - h) = \frac{9.1 + 5.5h}{2} (h) (219.5 + 57.9 + 12h)$$

$$(3{,}014 + 5{,}580 - 747h + 25h^2) (3.96 - h) (\tfrac{1}{2}) = (\tfrac{1}{2}) (9.1 + 5.5h) (h) (277.4 + 12h)$$

$$34{,}032 - 2{,}958h + 99h^2 - 8{,}594h + 747h^2 - 25h^3 = 2{,}524h + 1{,}526h^2 + 109h^2 + 66h^3$$

$$34{,}032 - 11{,}552h + 846h^2 - 25h^3 = 2{,}524h + 1{,}635h^2 + 66h^3$$

$$14{,}076h + 789h^3 + 91h^3 = 34{,}032$$

This equation can be solved for the height of the dikes above ground level (h) by trial and error, choosing a value of h and solving for the value to the right of the equal sign.

h	f(h)
1.5 m (5 ft)	23,196
2.4 m (8 ft)	39,585
2.1 m (7 ft)	33,882

Cut Volume = 6,500 m^3 (8,500 yd^3)

Fill Volume = 6,340 m^3 (8,300 yd^3)

If the equalization pond site is level, 1.8 m (6 ft) of pond excavation will provide the fill necessary to build the dikes 2.1 m (7 ft) above ground level, giving a total height of 3.9 m (13 ft).

Total Water Surface Area = $(70.1\text{ m})^2$ = 4,915 m^2 (52,900 ft^2)

= 0.49 ha (1.21 ac)

Total Pond Area = $(96.6\text{ m})^2$ = 9,340 m^2 (100,500 ft^2)

= 0.94 ha (2.31 ac)

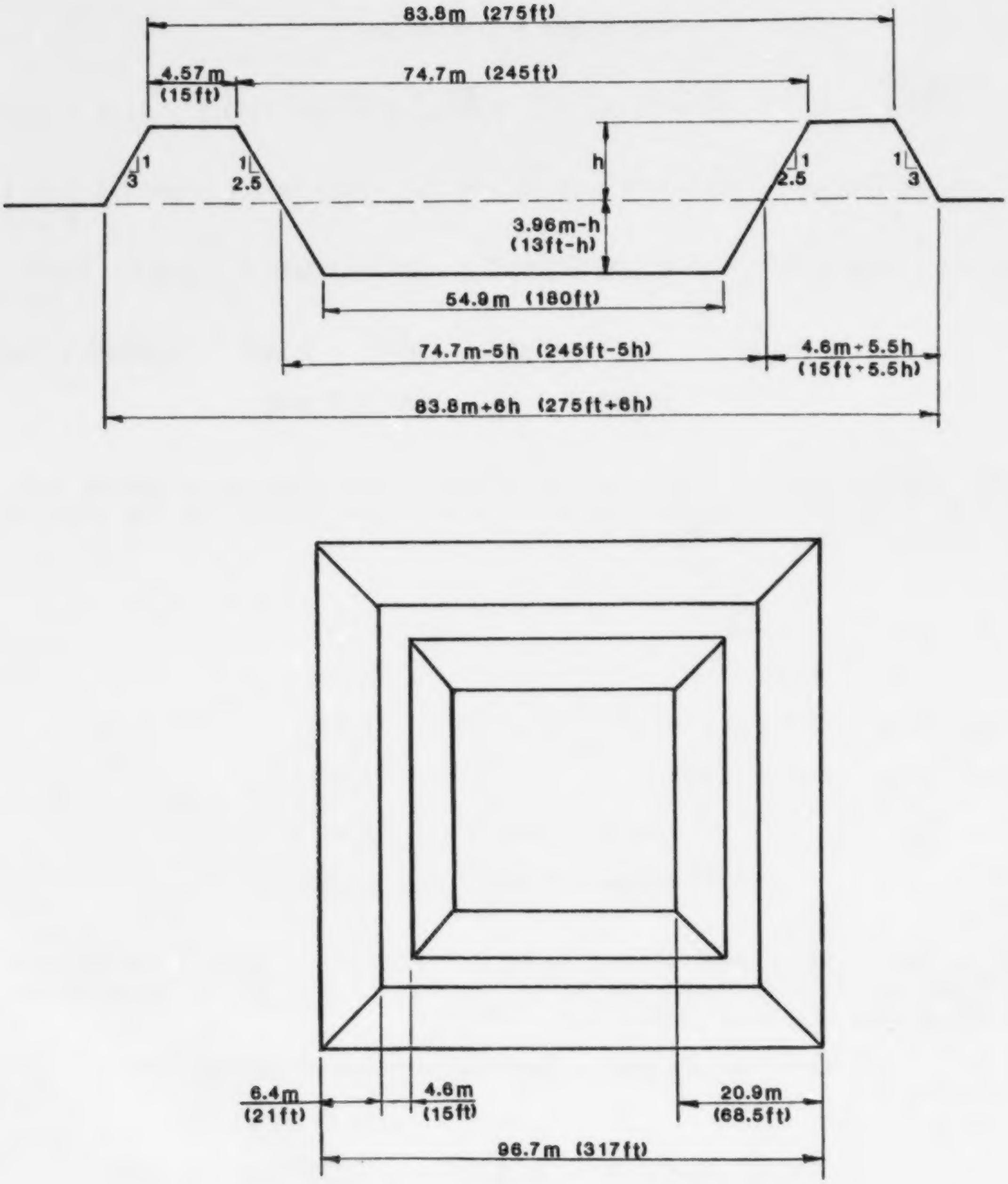
83.8m (275ft)
4.57m
(15ft)
74.7m (245ft)
h
1
3
1
2.5
1
2.5
1
3
3.96m-h
(13ft-h)
54.9m (180ft)
74.7m-5h (245ft-5h)
4.6m+5.5h
(15ft+5.5h)
83.8m+6h (275ft+6h)
6.4m
(21ft)
4.6m
(15ft)
20.9m
(68.5ft)
96.7m (317ft)

Figure 12-3. Equalization basin, Design Example II.

12.3.2 Theoretical Lime Requirement

Assume a 70% lime efficiency. The net alkalinity, as determined from analyses, is -2,200 mg/l as $CaCO_3$. The theoretical lime requirement from Table 1-2 is thus

$$2{,}200 \text{ mg} \times \frac{1 \text{ mol } Ca(OH)_2}{1.35 \text{ mol } CaCO_3} \times \frac{1}{0.7} = 2{,}325 \text{ mg/l } Ca(OH)_2$$

The theoretical daily lime ($Ca(OH)_2$) requirement is

$$\frac{2{,}325 \text{ mg}}{\text{l}} \times \frac{11{,}520 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = 26{,}784 \text{ kg/d } (58{,}925 \text{ lb/d}), \text{ or}$$

$$= 26.8 \text{ kkg/d } (29.5 \text{ tons/d})$$

From a titration curve, it was found that 12% excess lime is needed to maintain a pH of 9.0 for manganese removal, but 15% will be used in this calculation. The theoretical daily lime ($Ca(OH)_2$) requirement is then

$$26{,}784 \text{ kg/d} \times 1.15 = 30{,}802 \text{ kg/d } (67{,}764 \text{ lb/d}), \text{ or}$$

$$= 30.8 \text{ kkg/d } (33.9 \text{ tons/d})$$

The above procedure can be used as a preliminary estimation of the amount of lime that must be handled. It is a fairly accurate method, assuming the raw water quality will not change significantly. This method of approximating lime usage is the best alternative when a treatability test cannot be performed. When designing a plant of this size, however, a treatability test should be performed.

12.3.3 Actual Lime Requirement from Treatability Test

A treatability test was conducted according to the procedures outlined in Chapter 6, Section 6.3. This test indicated that 2,010 mg/l of lime are required to neutralize to pH 9.0. Therefore, the actual lime requirement, assuming constant water quality, is

$$\frac{2{,}010 \text{ mg}}{\text{l}} \times \frac{11{,}520 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg}} = 23{,}200 \text{ kg/d } (51{,}000 \text{ lb/d})$$

$$= 16.1 \text{ kg/min } (35.5 \text{ lb/min})$$

$$= 23.2 \text{ kkg/d } (25.6 \text{ tons/d})$$

At this point, the designer must consider the type of lime to be used for neutralization. In the past, the obvious choice would have been to use quicklime, because a significant price difference existed between quicklime and hydrated lime. The savings that resulted from using quicklime eventually offset the initial capital investment and maintenance costs associated with a slaker. The difference in cost between hydrated lime and quicklime has recently narrowed to within $2.20/kkg ($2.00/ton), making the choice less clear.

In this case, since the hydrated lime requirement is over 23 kkg/d (25 tons/d), a $50/d savings can be realized by installing a slaker and using quicklime. This is a significant savings that will offset the slaker capital cost and any anticipated maintenance costs. Therefore, the quicklime system is chosen.

12.3.4 Quicklime Requirements

It is important to realize that a slaker has an average efficiency near 90%. The remaining 10% is lost to grit, and lime leaving with the grit. Many vendors claim higher efficiencies; these claims may be true, making our value conservative. Slaker efficiency also depends on the quality of lime used.

The quicklime (CaO) requirement is calculated as follows:

$$\text{CaO Equivalent} = 23.2 \text{ kkg Ca(OH)}_2\text{/d} \times \frac{56 \text{ g CaO/mol}}{74 \text{ g Ca(OH)}_2\text{/mol}}$$

$$= 17.6 \text{ kkg CaO/d (19.4 tons/d)}$$

$$\text{Actual quicklime requirement} = \frac{17.6 \text{ kkg CaO/d}}{0.9} = 19.6 \text{ kkg CaO/d (21.6 tons/d)}$$

12.3.5 Lime Silo

The lime silo should be sized to provide at least 7 days storage.

$$\text{Silo Capacity} = 7 \text{ d} \times 19.6 \text{ kkg/d} = 137.2 \text{ kkg (151 tons)}$$

Assuming a quicklime density of 882 kg/m^3 (55 lb/ft^3), the silo volume required for 7 days quicklime storage is

$$\text{Silo Volume} = \frac{137{,}200 \text{ kg}}{882 \text{ kg/m}^3} = 155.6 \text{ m}^3 \text{ (5,500 ft}^3\text{)}$$

Silo Specifications:

diameter = 3.65 m (12 ft)

sidewall height = 13.7 m (45 ft)

60° hopper bottom

12.3.5.1 Bin Activator

A bin activator is highly recommended for silos storing quicklime. Many types of hoppers or bin activators are on the market. This designer chooses a 1.83-m (6-ft) vibratory bin activator, one-half the silo diameter.

12.3.5.2 Silo Slide Valve

This simple valve is considered mandatory on every lime storage silo. The slide valve fits over the hopper opening during filling to prevent lime from being blown into the feeder room below the silo.

12.3.6 Quicklime Feeder (to Slaker)

The required feeder rate is

$$\frac{19{,}600 \text{ kg/d}}{24 \text{ hr/d}} = 820 \text{ kg/hr } (1{,}800 \text{ lb/hr})$$

$$= 0.94 \text{ m}^3\text{/hr } (32.3 \text{ ft}^3\text{/hr})$$

A variable screw feeder is selected with an operating range of 0.56-5.60 m^3/hr (20-200 ft^3/hr), having a 0.56-kW (0.75-hp) motor.

12.3.7 Slaker

An automatic, thermostatically controlled, continuous slaker is chosen to maintain a slaking temperature of 77°C (170°F), and to provide an efficient and safer method of slaking. A slaker with a maximum rated capacity of 1,135 kg/hr (2,500 lb/hr) will meet the slaking needs of this plant.

12.3.8 Slurry Feed System

The operational criteria for the slurry feed system are listed below.

1. Use a loop system with a pH-controlled pinch valve. A 0.75-kW (1-hp) compressor, providing 5-atm (60-lb/in^2) pressure, is needed to operate the pinch valve. A pH control system will regulate the pinch valve.

2. Slurry addition should not exceed 10% of the raw water flow.

3. A minimum velocity of 1.22 m/s (4 ft/s) should be maintained in the slurry loop.
4. Circulate slurry at three times the average bleed-off rate.
5. Assume the slurry loop total length is 15.25 m (50 ft), the static head is 3 m (10 ft), and the friction head loss is 3 m (10 ft).

12.3.8.1 Evaluating Possible Slurry Concentrations

The slurry system is to deliver 16.28 kg/min (35.9 lb/min) $Ca(OH)_2$. Slurry concentrations are examined in Table 3-1.

The final choice of slurry concentration will involve a tradeoff between higher maintenance costs associated with high slurry concentrations and the resulting lower pump and power costs. This designer will use a 15% slurry in the loop system.

In this case the sulfate concentration (2,000 mg/l) in the raw AMD is safely below the concentration (3,000 mg/l) where gypsum formation will occur, so raw AMD can be used as makeup water. Where gypsum will be a problem, a separate water source such as a well is recommended.

12.3.9 Stabilization Tank

The tank will be designed to utilize the upper one-half of the tank volume. This will allow slurry storage for peak demand if the raw water requires more lime. Even at the low level point, the tank should provide 20 minutes detention time for the incoming slurry.

One-half tank volume = 20 min x 1.8 l/s x 60 s/min = 2,200 l (581 gal)

Provide a tank twice this size to provide 20 minutes detention when the tank is half full.

$$\text{Tank volume} = 4{,}400 \text{ l } (1{,}162 \text{ gal})$$

$$= 4.40 \text{ m}^3 \ (155.4 \text{ ft}^3)$$

Design a circular tank with the diameter equal to the height. Assume H=D.

$$\text{Vol} = 0.785\ (D^2)\ (H)$$

$$4.40 = 0.785\ (D)^3$$

$$D = 1.78 \text{ m } (5.8 \text{ ft})$$

The final tank dimensions are as follows:

Diameter = 1.83 m (6 ft)

Height = 2.44 m (8 ft), which includes 0.61 m (2 ft) freeboard

Volume = 4.8 m^3 (1,269 gal)

12.3.9.1 Level Controls

Level controls should be placed in the stabilization tank as follows:

1. high-level emergency shutoff at 2.13 m (7 ft);
2. high-level shutoff at 1.83 m (6 ft);
3. low-level slaker starting switch at 0.91 m (3 ft).

12.3.9.2 Miscellaneous Tank Requirements

The following are also included in the design:

1. full drain bottom nozzle;
2. elevated tank to provide flooded suction;
3. a centrifugal slurry pump and a standby slurry pump piped in parallel.

12.3.9.3 Slurry Tank Mixer

The previous rule of thumb, 0.2 kW/m^3 (1 hp/1,000 gal), will give a good approximation of the required mixer size. Therefore, a 4,800 l (1,269 gal) tank requires a 0.95-kW (1.27-hp) mixer. Use a 1.1-kW (1.5-hp) propeller mixer with 1.5 pitch at 350 r/min. The designer has the option to side-mount or fix-mount the mixer on a small cross beam at a slight angle. This would eliminate the need for a baffled slurry tank. The designer can also choose a top-mounted, vertical, on-center mixer with a baffled tank.

12.3.10 Flash Mix Tank

The flash mix tank should provide approximately 5 minutes detention time for neutralization. The required volume is

$$\text{Volume} = Q(\text{flow}) \times t\ (\text{time})$$

$$= (133.3 + 1.8\ \text{l/s}) \times (5\ \text{min})\ (60\ \text{s/min})$$

$$= 40{,}550\ \text{l}\ (10{,}713\ \text{gal})$$

$$\text{Volume} = 40.55\ \text{m}^3\ (1{,}432\ \text{ft}^3)$$

Design a circular tank with the diameter equal to the height.

$$\text{Volume} = 0.785\ D^2\ (H) \qquad \text{Set } D = H$$

$$40.55\ m^3 = 0.785\ D^3$$

$$D^3 = 51.7\ m^3$$

$$D = 3.72\ m\ (12.22\ ft)$$

Make the diameter 3.81 m (12.5 ft). The final tank dimensions are

Diameter = 3.81 m (12.5 ft)

Height = 4.42 m (14.5 ft), which includes 0.61 m (2 ft) freeboard

12.3.10.1 Baffles

Four baffles should be placed in the tank, 90° apart. Design specifications are

Width = (3.81 m) x (1/12) = 0.32 m (12.5 in)

Wall clearance = 0.15 x (0.32 m) = 0.048 m (2 in)

Bottom clearance = 0.5 (0.32 m) = 0.16 m (6 in)

Extended baffles = 0.30 m (1 ft) above static water level

12.3.10.2 Mixer

Using 0.2 kW/1 m^3 (1 hp/1,000 gal),

$$40.55\ m^3 \times 0.2\ kW/1\ m^3 = 8.0\ kW\ (10.7\ hp)$$

A 7.46-kW (10-hp) mixer will be satisfactory. It should be a top-mounted, vertical, on-center, axial turbine mixer. The influent pipes should enter the top with an air break, while the outlet should be an inverted elbow at 3.81 m (12.5 ft) with a 0.91-m (3-ft) submerged nipple.

The inlet and outlet must be 180° apart.

12.3.11 Aeration Tank

The theoretical iron loading is as follows:

$$\frac{440\ \text{mg}}{1} \times \frac{11,520\ \text{m}}{\text{d}} \times \frac{1,000\ 1}{\text{m}^3} \times \frac{1\ \text{kg}}{10^6\ \text{mg}} = 5,069\ \text{kg/d}\ (11,165\ \text{lb/d})$$

The theoretical oxygen requirement per day is

$$\frac{5,069\ \text{kg}}{7} = 724\ \text{kg/d}\ (1,595\ \text{lb/d})$$

A mechanical or surface aerator is selected for use. Based on an oxygen transfer rate of 2.13 kg/kW-hr (3.5 lb/hp-hr), an aerator of approximately 14.9 kW (20 hp) is needed. The mixer must be checked after the aeration tank is sized to insure complete mixing.

12.3.11.1 Aeration Detention Time

From the treatability study, an iron oxidation rate of 15 mg/l/min was determined. Based upon this rate, a detention time is calculated, assuming total oxidation.

$$D_t = \frac{440\ \text{mg/l}}{15\ \text{mg/l/min}} = 29\ \text{min}$$

Employing a safety factor of two, which takes into consideration lower oxidation rates from weather, short-circuiting, higher flows, and an increase in iron concentrations, a total detention time of 58 minutes should be provided.

The aeration basin volume required is

Volume = flow x time

= (133.3 + 1.8 l/s) (58 min) (60 s/min)

= 470,148 l (124,213 gal)

= 470.1 m^3 (16,606 ft^3)

Because the area available is limited, the aeration basin will be constructed of either steel or concrete. Also, it will be circular.

Volume = 0.785 D^2H Assume: H = 4.6 m (15 ft)

470.1 m^3 = 0.785 D^2 (4.6 m)

470.1 = 3.61 D^2

D^2 = 130 m^2

D = 11.4 m (37.4 ft)

The aeration tank dimensions are

Diameter = 12.2 m (40 ft)

Depth = 5.2 m (17 ft), which includes 0.61 m (2 ft) freeboard

It is imperative to insure that the aerator has enough power to turn over the basin contents. The geometry of the basin must also be within the aerator limits.

For the aerator selected, the specifications indicate it will provide a zone of complete mixing 17.6 m (58 ft) in diameter with a 4.5-m (15-ft) depth. This is adequate for this application and should provide a margin of safety.

12.3.12 Estimated Sludge Production

The sludge produced by the treatment of this water can be estimated as the sum of the metal hydroxides removed in significant concentrations, iron and aluminum, suspended solids in the mine drainage, and unused lime. This estimation assumes 100% removal.

Hydrolysis of ferric iron, assuming complete oxidation of ferrous iron:

$$\mathrm{Fe}^{+3} + 3\mathrm{H_2O} \rightarrow \mathrm{Fe(OH)_3} + 3\mathrm{H}^+$$

56 g/g-mol   107 g/g-mol

480 mg/l $\frac{107}{56}$ (480 mg/l) = 917 mg/l

Aluminum hydrolysis:

$$\mathrm{Al}^{+3} + 3\mathrm{H_2O} \rightarrow \mathrm{Al(OH)_3} + 3\mathrm{H}^+$$

27 g/g-mol   78 g/g-mol

150 mg/l $\frac{78}{27}$ (150 mg/l) = 433 mg/l

Theoretical daily solids production:

Ferric hydroxide

$$917 \text{ mg/l} \times \frac{11{,}520 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg/l}} = 10{,}560 \text{ kg/d } (23{,}270 \text{ lb/d})$$

$$433 \text{ mg/l} \times \frac{11{,}520 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg/l}} = 4{,}990 \text{ kg/d } (10{,}990 \text{ lb/d})$$

$$165 \text{ mg/l} \times \frac{11{,}520 \text{ m}^3}{\text{d}} \times \frac{1{,}000 \text{ l}}{\text{m}^3} \times \frac{1 \text{ kg}}{10^6 \text{ mg/l}} = 1{,}900 \text{ kg/d } (4{,}190 \text{ lb/d})$$

Unused lime (assume 10%)

$$23{,}200 \text{ kg/d} \times 0.10 = 2{,}320 \text{ kg/d } (5{,}100 \text{ lb/d})$$

Total solids, dry weight = 19,770 kg/d (43,550 lb/d)

Assuming the sludge solids content is 1%, the daily wet sludge production is

$$\text{Sludge} = \frac{19{,}770 \text{ kg/d}}{0.01}$$

$$= 1{,}977{,}000 \text{ kg/d } (4{,}355{,}000 \text{ lb/d})$$

Assuming the sludge weighs the same as water, the volume of sludge produced daily is

$$\text{Sludge Volume} = \frac{1{,}977{,}000 \text{ kg/d}}{1 \text{ kg/l}} = 1{,}977{,}000 \text{ l/d } (522{,}300 \text{ gal/d})$$

$$= 1{,}977 \text{ m}^3\text{/d } (69{,}800 \text{ ft}^3\text{/d})$$

12.3.13 Actual Sludge Production from Treatability Test

The above calculations could be used when a treatability test is not performed. The better way to design a plant of this size, however, is with data obtained from a treatability test.

The sludge volume based upon the treatability test amounted to 11% of the plant flow, 1,267 m^3/d (334,400 gal/d), which is lower than the theoretical value. This could be caused by several factors: (1) high suspended solids in the supernatant, whereas the theoretical value assumed removal to 35 mg/l; (2) better lime efficiency; and (3) ideal settling conditions producing a sludge density greater than 1%. Therefore, it is the designer's option as to which volume to use. This designer chooses to be conservative and use the larger volume, 1,267 m^3/d (334,400 gal/d).

12.3.14 Settling Unit

Because of limited land availability, a mechanical clarifier will be provided. Using a rise rate of 0.006 m/min (0.02 ft/min), the clarifier diameter is sized accordingly.

$$\text{Area} = \frac{\text{Flow}}{\text{Rise Rate}} = \frac{11{,}520 \text{ m}^3\text{/d}}{0.006 \text{ m/min} \times 1{,}440 \text{ min/d}} = 1{,}330 \text{ m}^2 \ (14{,}350 \text{ ft}^2)$$

$$\text{Area} = 0.785 D^2$$

$1{,}330\ \text{m}^3 = 0.785\ D^2$

$D = 41.2\ \text{m}\ (135\ \text{ft})$

The depth of the clarifier or thickener is a function of detention time, raking mechanism, mode of operation, and type of separator chosen (solids contact vs. conventional). The designer chooses an upflow clarifier with 2 days sludge storage capacity, and 12 hours clear water depth.

The clarifier volume is

2 days sludge storage = 2,530 m^3 (668,800 gal)

12 hours clear water depth = 5,760 m^3 (1,522,000 gal)

Total Volume = 8,290 m^3 (2,190,800 gal)

The clarifier depth (H) is

$\text{Volume} = 0.785\ (D)^2\ (H)$

$8{,}290\ \text{m}^3 = 0.785\ (41.2\ \text{m})^2\ (H)$

$H = 6.22\ \text{m}\ (20.4\ \text{ft})$

The final clarifier dimensions will be those of the closest standard-size clarifier.

12.3.15 Final Sludge Disposal

Final sludge disposal will be to a borehole.

CHAPTER 13

COST ESTIMATING

13.1 Introduction

Determining the capital investment and construction cost of any proposed AMD treatment plant, as well as its daily or annual cost of operation, is an important part of the designer's role.

There are many ways to estimate the cost of construction, but this will vary with the designer's capability. Most designers are able to perform the engineering and construction in-house, while others might hire a consultant to do the engineering design and then contract for the plant's construction. All of these factors influence the total cost of a facility and lead to large cost differences between similar plants.

Construction cost estimates can be prepared in different ways. For initial budget purposes, determining the installed or constructed costs for the various major units of a treatment facility by rule-of-thumb methods and adding in the costs for associated facilities or services will enable the designer to establish an initial budget within 25% of the final cost. This method of estimating can also be used for comparing costs among any options within a unit process. The cost items to be considered could include the following:

Major Cost Items	Associated Facilities
Borehole and pumps	Land and access roads
Equalization basin and pumps	Zoning and permits
Lime system	Power supply
Aeration system	Interconnecting piping
Settling unit(s)	Panels and wiring
Sludge dewatering	Instrumentation
Sludge disposal	Special foundations or site preparation
	Fencing

Contingency

Engineering fees

Escalation of cost by inflation during the construction period

Contractor's overhead and profit

Once an initial budget for the project is established, the cost estimate can be refined or finalized only by completing the engineering design necessary for construction of the facility and installation of its equipment. The degree of effort here will determine the accuracy of the cost estimate. Complete design and detail drawings will enable the estimator to refine his final budget cost to a contingency of less than 10%. In addition, this effort will also assist the contractors bidding for the work to provide more accurate prices.

For rapid estimating purposes, which will determine the "ballpark" cost of a treatment facility, Figure 13-1 can be used. This figure is based on the costs for many treatment facilities with varying levels of equipment and automation. A range of costs based on flow through the facility is shown by the band on the figure. This covers simple to complex levels of construction.

Typical initial budget cost estimates have been prepared for the two design examples presented in Chapter 12. These cost estimates include the rule-of-thumb methods discussed throughout this manual, as well as pricing values from the 1980 Dodge Manual for Building Construction Pricing and Scheduling and the 1980 Means Catalog. These are excellent sources for up-to-date construction estimating values.

Annual operational and maintenance costs, which include electricity and chemicals, are also included in the examples. They are based on 1981 product costs.

13.2 Cost Breakdown of Design Example I

This particular mine drainage plant has a daily average flow of 2,880 m^3/d (527 gal/min). Chemical requirements are 1.8 kkg/d (2 tons/d) of hydrated lime.

Capital Costs

The following capital costs will be incurred with this plant:

	Item	Cost ($)
1.	Raw Water Pumps (2) 500 gal/min	$ 2,000
2.	Equalization Basin, 6,000 m^3 (1.5 Mgal)	

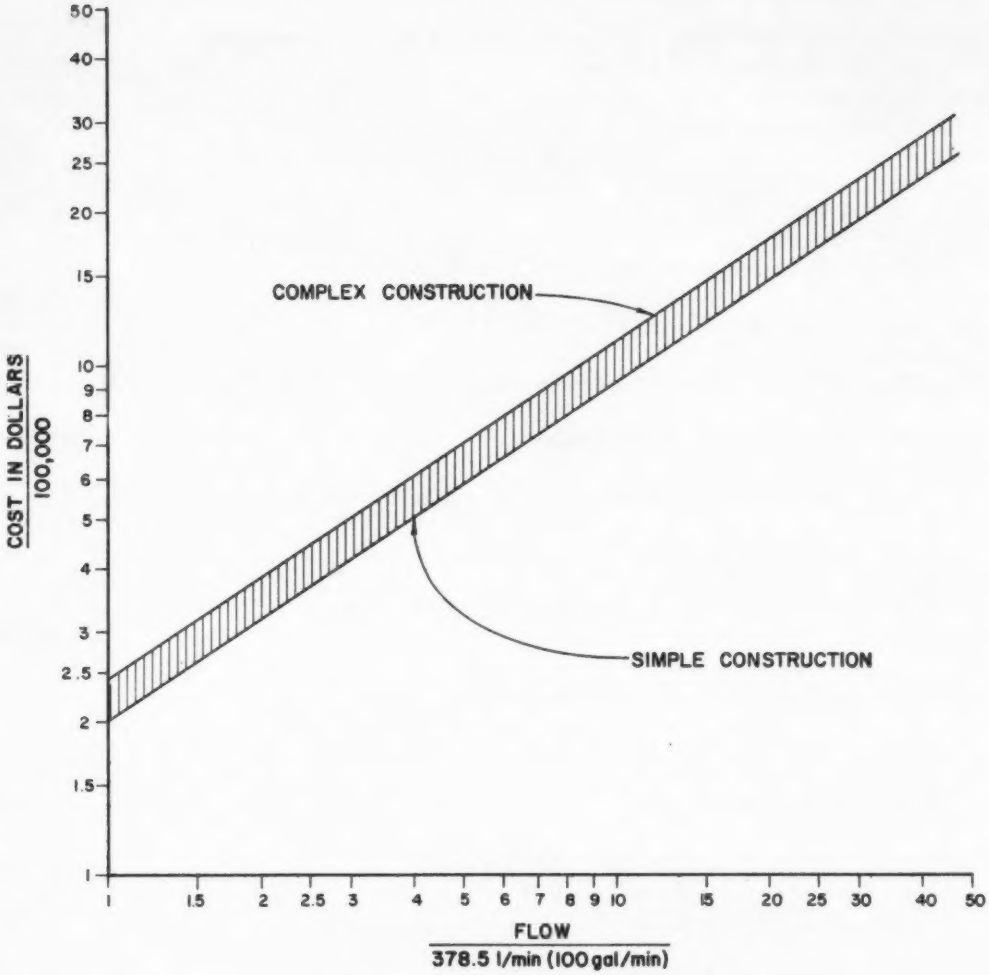

Figure 13-1. Installed cost vs. plant flow.

	Item	Cost ($)
	Excavation and material placement 6,000 m^3 (7,800 yd^3) @ $3/$yd^3$	$23,400
	Clay liner thickness 0.91 m (3 ft) Area equals 5,850 m^2 (63,000 ft^2) 5,324 m^3 (7,000 yd^3) @ $3/$yd^3$	21,000
3.	Lime Storage and Feed System	
	Silo 27.2 kkg (30 tons) including side ladder, handrail, cage, 60° hopper, fill pipe	10,000
	Bin activator 1.52 m (5 ft)	4,500
	5-cm (2-in) screw feeder	2,100
	Hopper slide valve	250
	Bin level indicator	250
	Dust collector	1,300
	Concrete foundation 3.65-m (12-ft) square pad x 1.06 m (3.5 ft) 14.5 m^3 (19 yd^3) @ $250/$yd^3$	4,750
	Pad excavation 29 m^3(38 yd^3) @ $4/$yd^3$	150
	Delivery	600
	Erection (minimal because welded silo)	560
4.	Lime Fed Dry Directly from Feeder	No Cost
5.	Flash Mix Tank approximately 11,355 l (3,000 gal)	
	Diameter = 2.44 m (8 ft) Height = 3.05 m (10 ft) Fiberglass tank includes baffles, nozzles, mixer mounts	4,100
6.	Aeration Tank (Circular Earthen Basin)	
	Diameter = 6.40 m (21 ft) Depth = 1.83 m (6 ft) 59 m^3 (77 yd^3) @ $3/$yd^3$	230
	Bottom concrete formless pour 5 cm (2 in) 37 m^2 (400 ft^2) x .05 m = 2.0 m^3 2.0 m^3 (2.6 yd^3) @ $200/$yd^3$	525
7.	Aerator 0.89 kW (1.2 hp)	3,500
8.	Settling Basin (Earthen) with Sludge Removal System	
	Volume 1,782 m^3 (471,000 gal)	
	From cut and fill Fill + excavation = 2,910 m^3 (3,810 yd^3) @ $3/$yd^3$	11,430

	Item	Cost ($)
	Clay liner 0.61 m (2 ft) x 2,809 m^2 (30,240 ft^2) 1,713 m^3 (2,240 yd^3) @ $\$3/\text{yd}^3$	6,720
	Hydraulic sludge removal system	15,000
9.	Sludge Disposal Pond 46,000 m^3 (12.15 Mgal)	
	From cut and fill 97,379 m^3 (124,254 yd^3) @ $\$3/\text{yd}^3$	372,764
	Clay liner 0.61 m (2 ft) x 28,200 m^2 (303,546 ft^2) 17,202 m^3 (22,500 yd^3)@ $\$3/\text{yd}^3$	67,498
10.	Land Cost for Treatment Site (assume 62.5 ha)	
	6 ac @ $5,000/ac	30,000
11.	Instrumentation for Automatic Interlock System with pH Assemblies, Panel, Annunciator, and Recorder	5,000
12.	Electrical Motor Starts, Transformer, Heater, Control Room with Insulation, Lighting	5,000
13.	Piping and Miscellaneous	5,000
14.	Electric Power to the Site (assume $25,000)	25,000
15.	Fencing, Complete Enclosure	
	$33.00/m ($10/ft) x 3,100 m	31,000
	Initial Construction Cost Estimate	$653,627
	Engineering Fees @ 10%	65,363
	Contingency @ 15%	98,050
	Contractor's Overhead and Profit @ 20%	150,335
	TOTAL CAPITAL COST BUDGET	$967,375

Annual Operational and Maintenance Costs

1.	Electricity (11.7 KwH @ $0.05/KwH)	$ 5,130
2.	Chemicals (lime @ $65/ton)	47,450

	Item	Cost ($)
3.	Manpower (1 worker @ 4 hr/d @ $10/hr)	14,600
4.	Sludge Disposal (assume $10,000/yr for maintenance of hydraulic sludge removal system)	10,000
		$77,180

13.3 Cost Breakdown of Design Example II

This particular mine drainage plant has a daily average flow of 11,520 m^3 (2,110 gal/min). Chemical requirements are 17.6 kkg (19.4 tons) of quicklime per day.

Capital Costs

	Item	Cost ($)
1.	Raw Water Pumps (2)	$ 4,500
2.	Equalization Basin 11,520 m^3 (3.0 mg)	
	Cut and fill 12,840 m^3 (16,795 yd^3) @ $3/yd^3	50,385
	Clay liner 0.91 m (3 ft) 9,336 m^2 (100,500 ft^2) @ 0.91 m (3 ft) = 8,496 m^3 (11,113 yd^3) @ $3/yd^3	33,340
3.	Lime Storage Silo Capacity 137 kkg (151 tons)	
	This will be a completely welded tank including caged side ladder, handrail, 60° hopper, fill pipe	21,000
	Bin activator 1.82 m (6 ft)	5,000
	10.2-cm (4-in) screw feeder	3,200
	Hopper slide valve	350
	Bin level indicators	250
	Dust collector	2,600
	Concrete foundation 4.6-m (15-ft) square pad x 1.5 m (5 ft) 32 m^3 (42 yd^3) @ $250/yd^3	10,500
	Pad excavation 64 m^3 (84 yd^3) @ $4/yd^3	328
	Delivery	1,200
	Erection (welded silo)	1,120

	Item	Cost ($)
4.	Lime Feed System (Slaking Quicklime)	
	1,135 kg/hr (2,500 lb/hr) slaker	15,200
	Slaker installation	2,500
	Slurry loop (piping and pinch valve)	1,000
	Compressor 0.75 kW (1 hp) and installation	1,200
	Recirculation pump 0.37 kW (0.5 hp)	750
	Stabilization tank 4,400 l (1,162 gal) fiberglass material, mixer mounts, level controls, nozzles	2,100
	Propeller mixer 1.10 kW (1.5 hp) stainless steel shaft and propeller	1,500
	Miscellaneous fabrication and piping	1,000
5.	Flash Mix Tank 50.5 m^3 (13,500 gal)	
	Coated steel tank, baffles, nozzles, mixer mounts	5,500
	Mixer 7.46 kW (10 hp)	8,000
6.	Circular Concrete Aeration Basin	
	604.5 m^3 (160,000 gal) volume	
	Excavation 1,224 m^3 (1,600 yd^3) @ \$3/yd^3	4,800
	Concrete for walls 78 m^3 (102 yd^3) @ \$250/yd^3	25,500
	Concrete for pad 135 m^3 (177 yd^3) @ \$250/yd^3	44,250
	Aerator 14.9 kW (20 hp) fix-mounted	22,000
	Aerator walk-on platform	5,000
7.	Settling Unit (Mechanical Clarifier)	
	Concrete structure with steel mechanism Single unit 41.2 m (135 ft) in diameter Volume is 8,290 m (2,190,300 gal) Clarifier will be half-buried Excavation 6,373 m^3 (8,335 yd^3) @ \$3/yd^3	25,000
	Clarifier wall concrete 463 m^3 (605 yd^3) @ \$175/yd^3	105,875
	Bottom structure and pad 1,737 m^3 (2,272 yd^3) @ \$175/yd^3	397,600
	Mechanism 41.2 m (135 ft) @ \$1,250/ft	168,750
	Piping and miscellaneous	10,000
8.	Sludge Disposal to Borehole	Variable
9.	Land Cost for Treatment Site (assume 2 ha)	
	5 ac @ \$5,000	25,000

	Item	Cost ($)
10.	Instrumentation for Automatic Interlock System, pH Assembly, Pumps, Control Panel, Annunciator, Recorder, and Housing	30,000
11.	Electrical Motor Starters, Transformer Heater, Lighting, Insulation, Wiring	30,000
12.	Piping and Miscellaneous	15,000
13.	Electrical Power to Site	Variable
14.	Fencing Minimal	$ 15,000
	Initial Construction Cost Estimate	$1,096,298
	Engineering Fees @ 10%	109,630
	Contingency @ 15%	164,445
	Contractor's Overhead and Profit @ 20%	252,148
	TOTAL COST BUDGET	$1,622,521

Annual Operational and Maintenance Costs

1.	Electricity (36.5 KwH @ $0.05/KwH)	$ 16,000
2.	Chemicals (19.4 tons quicklime/d @ $60/ton)	424,860
3.	Manpower (1 worker @ 6 hr/d @ $10/hr)	21,900
		$ 462,760

APPENDIX

PHYSICAL AND CHEMICAL PROPERTIES OF LIME

A.1 Specifications on Lime

Use of the following standards by the American Society for Testing and Materials specifying the nomenclature and the chemical and physical methods of testing chemical lime products is recommended.

C 51-71 - Terms Relating to Lime

C 50-57 - Sampling, Inspection, Packing, and Marking of Quicklime and Lime Products

C 110-71 - Physical Testing of Quick and Hydrated Lime

C 400-64 - Testing Quicklime and Hydrated Lime for Neutralization of Waste Acid

C 25-72 - Chemical Analysis of Limestone, Quicklime, and Hydrated Lime

C 53-63 - Quicklime and Hydrated Lime for Water Treatment

C 433-63 - Quicklime and Hydrated Lime for Hypochlorite Bleach Manufacture

C 415-72 - Quicklime and Hydrated Lime for Calcium Silicate Products

C 258-52 - Quicklime for Calcium Carbide Manufacture

C 259-52 - Hydrated Lime for Grease Manufacture

C 49-57 - Quicklime and Hydrated Lime for Silica Brick Manufacture

C 46-62 - Quicklime and Limestone for Sulfite Pulp Manufacture

C 45-25 - Quicklime and Hydrated Lime for Cooking of Rags in Paper Manufacture

C 593-69 - Fly Ash and Other Pozzolans for Use with Lime

Copies of these standards may be obtained by writing to the American Society for Testing and Materials, 1916 Race Street, Philadelphia, Pa., 19103.

TABLE A-1

TYPICAL ANALYSES OF COMMERCIAL QUICKLIMES

Component	High Calcium Quicklimes Range[a] %	Dolomitic Quicklimes Range[a] %
CaO	93.25 - 98.00	55.50 - 57.50
MgO	0.30 - 2.50	37.60 - 40.80
SiO_2	0.20 - 1.50	0.10 - 1.50
Fe_2O_3	0.10 - 0.40	0.05 - 0.40
Al_2O_3	0.10 - 0.50	0.05 - 0.50
H_2O	0.10 - 0.90	0.10 - 0.90
CO_2	0.40 - 1.50	0.40 - 1.50

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[a] The values given in this range do not necessarily represent minimum and maximum percentages.

TABLE A-2

pH OF CALCIUM HYDROXIDE SOLUTIONS AT 25°C

CaO g/l	pH
0.064	11.27
0.065	11.28
0.122	11.54
0.164	11.66
0.271	11.89
0.462	12.10
0.680	12.29
0.710	12.31
0.975	12.44
1.027	12.47
1.160	12.53

Since solubility of lime decreases as the temperature increases, the pH of lime solutions is correspondingly lower at higher temperatures.

Data from F.M. Lea and G.E. Bessey, Journal of the Chemical Society, 1,612-1,615, 1937.

TABLE A-3

PROPERTIES OF THEORETICALLY PURE LIME COMPONENTS

	Pure Lime			
Properties	Quicklime Components		Hydrated Lime Components	
Chemical Name	Calcium Oxide	Magnesium Oxide	Calcium Hydroxide	Magnesium Hydroxide
Chemical Formula	CaO	MgO	$Ca(OH)_2$	$Mg(OH)_2$
Molecular Weight	56.08	40.32	74.096	58.336
Melting Point	2,570°C (4,658°F)	2,800°C (5,072°F)	--	--
Decomposition Point	--	--	580°C (1,076°F)	345°C (653°F)[a]
Boiling Point	2,850°C (5,162°F)	3,600°C (6,512°F)	--	--
Refractive Index	1.838	1.736	1.574 and 1.545	1.559 and 1.580
Heat of Solution at 18°C	+18.33 kg-cal	--	+2.79 kg-cal	-0.0 kg-cal
Crystalline Form	cubic	cubic	hexagonal	hexagonal
Density	3.40	3.65	2.343	2.4
Solubility: In Hot and Cold Water	--	--	See Solubility Sections A.2 and A.4.	

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[a] There is not complete agreement on the exact decomposition point of $Mg(OH)_2$; however, the value given represents the best data available.

TABLE A-4

GRAVIMETRIC PERCENTAGES OF CRITICAL CONSTITUENTS OF LIMES

Lime Substances	Percents of Elements					Percents of Compounds		
	Ca	Mg	O	H	H_2O	CaO	MgO	Alkali Oxides (CaO + MgO)
CaO	71.47	--	28.53	--	--	100.00	--	100.00
$Ca(OH)_2$	54.09	--	43.19	2.72	24.31	75.69	--	75.69
MgO	--	60.32	39.68	--	--	--	100.00	100.00
$Mg(OH)_2$	--	41.69	54.85	3.46	30.88	--	69.12	69.12
CaO·MgO	41.58	25.23	33.19	--	--	58.17	41.83	100.00
$Ca(OH)_2 \cdot MgO$	35.03	21.26	41.95	1.76	15.75	49.01	35.24	84.25
$Ca(OH)_2 \cdot Mg(OH)_2$	30.27	18.36	48.33	3.04	27.21	42.35	30.44	72.79

TABLE A-5

PROPERTIES OF TYPICAL COMMERCIAL LIME PRODUCTS

QUICKLIMES

	High Calcium	Dolomitic
Primary Constituents	CaO	CaO and MgO
Specific Gravity	3.2 - 3.4	3.2 - 3.4
Bulk Density (Pebble Lime), lb/ft^3	55 - 60	55 - 60
Specific Heat at 38°C (100°F), BTU/lb. . .	0.19	0.21
Angle of Repose.	55°[a]	55°[a]

HYDRATES

	High Calcium	Normal Dolomitic	Pressure Dolomitic
Primary Constituents . .	$Ca(OH)_2$	$Ca(OH)_2 + MgO$	$Ca(OH)_2 + Mg(OH)_2$
Specific Gravity	2.3 - 2.4	2.7 - 2.9	2.4 - 2.6
Bulk Density, lb/ft^3 . .	25 - 35[b]	25 - 35[b]	30 - 40[b]
Specific Heat at 38°C (100°F) BTU/lb	0.29	0.29	0.29
Angle of Repose.	70°[a]	70°[a]	70°[a]

---

[a] The angle of repose for both types of lime (hydrate in particular) varies considerably with mesh, moisture content, degree of aeration, and physical characteristics of the lime (e.g., for quicklime it generally varies from 50°-55°, and for hydrated lime it may range as much as 15°-80°.

[b] In some instances, these values may be extended. The Scott method is used for determining the bulk density values. In calculating bin volumes, the lower figure should be used.

A.2 Solubility of Calcium Hydroxide

TABLE A-6

SOLUBILITY OF CALCIUM HYDROXIDE IN WATER

t°C	Grams/100 g Saturated Solution	
	CaO	$Ca(OH)_2$
0	0.140	0.185
10	0.133	0.176
20	0.125	0.165
25	0.120	0.159
30	0.116	0.153
40	0.106	0.140
50	0.097	0.128
60	0.088	0.116
70	0.079	0.104
80	0.070	0.092
90	0.061	0.081
100	0.054	0.071

The solubility of commercial limes in water does not vary more than 7% from the solubility of pure calcium hydroxide. The differences are probably due to the trace amounts of sodium and potassium hydroxide in commercial limes. Magnesia, silica, and carbonate have no effect upon the solubility of ordinary lime, but may have a marked effect upon its rate of solution.

Particle size has considerable influence upon solubility. Freshly slaked lime, which is of small particle size, is about 10% more soluble than coarse particle or aged slake lime. This effect is due to the slow expansion of the dry lime particles during storage.

These solubility data are derived from A. Seidell, Solubilities of Inorganic and Metal Organic Compounds, 3rd ed., vol. 1, 209-210, 1940.

A.3 Calculating Weights of Slurry

For calculating the weights of slurry with varying percentages of water, the following formula may be used:

$$W = \frac{6{,}237s}{100 - a + sa}$$

where W = weight in pounds of slurry per cubic foot

s = specific gravity of dry lime solids

a = percent water in slurry

A.4 Solubility of Magnesium Hydroxide

Magnesium hydroxide is virtually insoluble in water. At 18°C (64°F) and 100° C (212°F), the solubilities are 0.0098 and 0.0042 g of Mg(OH) /l respectively, in a saturated solution. The presence of small quantities of NaCl and Na SO in the aqueous solution will increase the solubility of Mg(OH) slightly.

These solubility data are from A. Seidell, Solubilities of Inorganic and Metal Organic Compounds, 3rd ed., vol. 1, 982, 1940.

A.5 Heats of Reaction at 25°C

Hydration or Slaking

$CaO + H_2O = Ca(OH)_2$ - heat evolved = 15,300 cal/g mol
= 27,000 BTU/lb mol

$MgO + H_2O = Mg(OH)_2$ - heat evolved = 8,800 to 10,000 cal/g mol
= 14,400 to 18,000 BTU/lb mol

Carbonation

$CaO + CO_2 = CaCO_3$ - heat evolved = 43,300 cal/g mol
= 78,000 BTU/lb mol

$MgO + CO_2 = MgCO_3$ - heat evolved = 28,900 cal/g mol
= 52,000 BTU/lb mol

Derived from Int. Crit. Tables, vol. V, 195-196.

* U.S. GOVERNMENT PRINTING OFFICE: 1983-659-095/0571